THERMAL PROCESSING OF
UTAH TAR SANDS

by

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SUPERVISORY COMMITTEE APPROVAL

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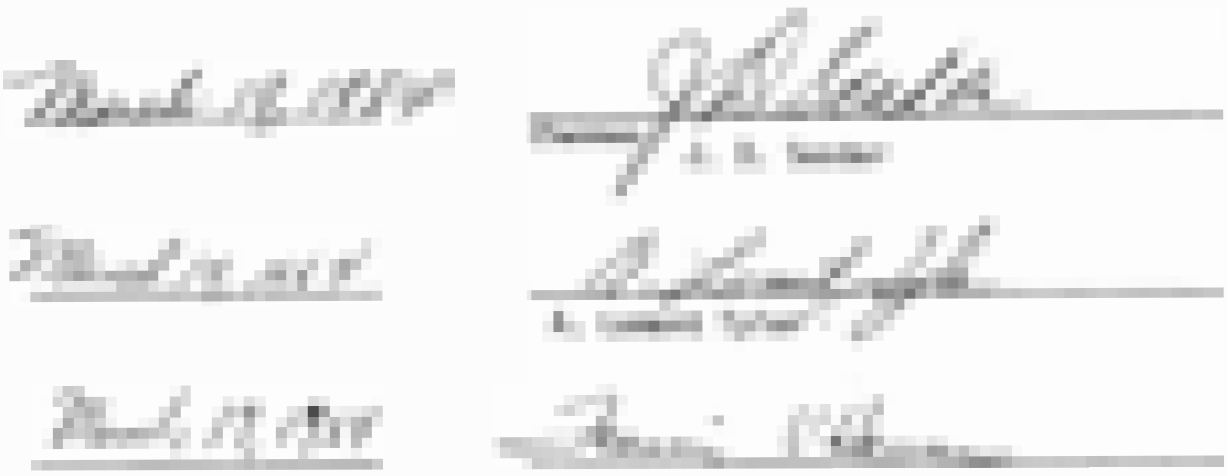

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ABSTRACT

A two-staged fluidized-bed reactor for the energy-efficient, thermal recovery of bitumen from Utah tar sands has been constructed. This reactor is a scaled-up version of an earlier system investigated at the University of Utah, and involves the use of three liquid-potassium heat pipes which thermally couple an upper pyrolysis bed with a lower combustion bed. The reactor has been studied to determine the effect of multiple heat pipes, increased feed rate, and longer duration run times.

The process consists essentially of three steps. In the first step, mined and suitably sized tar sand, being fed into the reactor at a constant rate, is pyrolyzed at temperatures of 440°C to above 500°C in an inert atmosphere to volatilize and partially crack most of the contained bitumen. The vaporized products of the pyrolysis section are condensed and coalesced to give a synthetic crude oil. In the second step, coked sand, formed as a by-product in the pyrolysis reactor, is combusted with air at temperatures between 550 and 600°C. In the third step, heat is recovered from the hot spent sand leaving the combustion bed by a fluidized-bed heat exchanger using vertical copper cooling tubes.

Three heat pipes are used to transfer 100% of the energy required to maintain the required temperature of pyrolysis in the upper bed. This energy is provided by the lower, combustion bed. Air is used to fluidize the combustion bed and the heat recovery bed. The pyrolysis bed is fluidized with nitrogen. In a commercial unit, recycled gas from

pyrolysis could be used in place of nitrogen to fluidize the pyrolysis bed. The process is equipped with a digital control and data acquisition system to permit a more efficient method of process study. The current reactor is capable of processing over 30 lbs of tar sand per hour, and has successfully operated on a continual basis for over 6 hours.

Over 30 runs, with tar sand from the Sunnyside, PR Springs, and Whiterocks deposits, have been made to study controllability of the process, heat requirements, and liquid product yield with respect to process variables such as pyrolysis-bed temperature and average solids-residence times. The highest liquid yield obtained was 71 weight percent of the original tar sand bitumen. Typically, coke yield is from 15 to 20 weight percent, with gas yield making up the difference.

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CHAPTER I

OBJECTIVES

The objectives for this research were as follows:

1. Design a scaled-up reactor to process tar sand and recover synthetic liquid crude oil from the tar sand bitumen by using data obtained from a previously operated, two-staged fluidized-bed thermal reactor, which used a single liquid-potassium heat pipe to thermally couple the two fluidized beds.

2. Construct the reactor using suitable materials.

3. Assemble the processing unit and equip it with the necessary instrumentation to control the process and acquire pertinent data.

4. Test the assembled reactors in the areas of thermal performance, solids handling, and controllability.

5. Process various Utah tar sands to study the effects of increased processing capacity and multiple heat pipes.

6. Identify possible optimizing trends relating liquid crude oil yield and different processing variables, such as average solids residence time and pyrolysis bed temperature.

7. Analyze the products produced during operation of the apparatus to assess their value as possible refinery feed stocks.

CHAPTER II

BACKGROUND

Tar Sand Resources

Tar sands, biomass, heavy oil, oil shale, and coal are considered to be the primary contributors to the emerging synfuel industry. On a global basis, Meyer and Fulton (1982) estimate the total heavy oil and bitumen reserves to be more than 4 trillion barrels with an in-place resource estimated at 671 billion barrels of oil, reserves being defined as the amount of in-place resource that can be recovered using available technology. These resources, which are enormous when compared to the known conventional light oil reserves, have recently engendered a continual growth of interest as can be illustrated by the occurrence of several significant events since 1969.

First, there has been the formation of the heavy oil and tar sand information center in New York for the collection, collation, and dissemination of information concerning the development of this resource. Second, definitions have been developed for the standardization of information; for instance, heavy oil is defined as having a viscosity of between 100 and 10,000 centipoise or an API gravity between 10° and 20°, and bitumen is defined as having a viscosity greater than 10,000 centipoise or an API gravity of less than 10° (Meyer, et al., 1982).

Third, the U. S. Geological Survey has initiated a study of the heavy oil and bitumen deposits of the United States for the first time in 15 years. And fourth, there has been an unceasing decline of the

known reserves of crude oil having an API gravity greater than 20°, accompanied by high crude oil prices, advances in upgrading technology, and an increasing desire for energy self-sufficiency.

Bitumen deposits represent about 70 percent of the total world resource of bitumen and heavy oil. Approximately 71 percent of the total heavy oil resource is believed to be located around the Orinoco Heavy Oil Belt in Eastern Venezuela; whereas, one of the latest estimates by Meyer, et al., (1982) places nearly 68 percent of the world bitumen resource in the western provinces of Canada. Listed in Table 1 is a summary of the world bitumen resources and reserves (Meyer et al., 1982).

The bitumen resource exists usually in the form of oil-impregnated rock deposits which are also referred to as tar sand, oil sand, or bituminous sands. Although Canada and the Soviet Union possess nearly ninety percent of the global tar sand resource, the United States' portion represents a significant fossil-fuel source. Even though tar sand deposits are known to exist in 22 of the 50 states, major deposits of 1 million barrels of in-place oil or more exist in only 9 states as is shown in Table 2.

Of the total United States tar sand resource, over 90 percent is estimated to be located in the state of Utah (Ritzma et al., 1982). Table 3 lists the major deposits of Utah, (Meyer, et al., 1982). The locations of the six giant deposits of Utah, which contain over 96 percent of the total bitumen resource of 50 Utah deposits, are shown in Fig. 1 (Ritzma, 1973).

Table 1

World Bitumen Summary
(million barrels)

Country	Oil in Place	Recoverable Reserves
Albania	371	56
Canada	2,432,600	253,970
Cuba		Trace
Italy	14,155	271
Madagascar	24,800	6,060
Mexico		19
Nigeria		5,022
Peru	66	13
Romania	25	4
Senegal		700
Trinidad	60	9
USSR	561,024	83,569
USA	34,890	15,840
Venezuela	12	70
West Germany		
Zaire	1,500	21
Total	3,069,002	347,624

Table 2

U.S. Bitumen Summary
(million barrels)

State	Original Oil in Place	Recoverable Reserves
Alabama	150	22
California	1,206	451
Kentucky	80	11
New Mexico	57	8
Ohio	1	Trace
Tri-State	287	43
Utah	29,608	14,655
Texas	2,000	500
Wyoming	1,000	150
Total	34,389	15,840

Table 3

Utah Tar Sand Summary
(millions of barrels)

Major Deposit	Estimated in-place oil	Recoverable Reserves
Asphalt Ridge	1,048	873
Asphalt Ridge Northwest	100	15
P.R. Spring	4,000	3,700
Hill Creek	1,160	830
Sunnyside	9,500	2,000
Tar Sand Triangle	12,500	6,100
Circle Cliffs	1,300	1,137
Totals	29,608	14,655

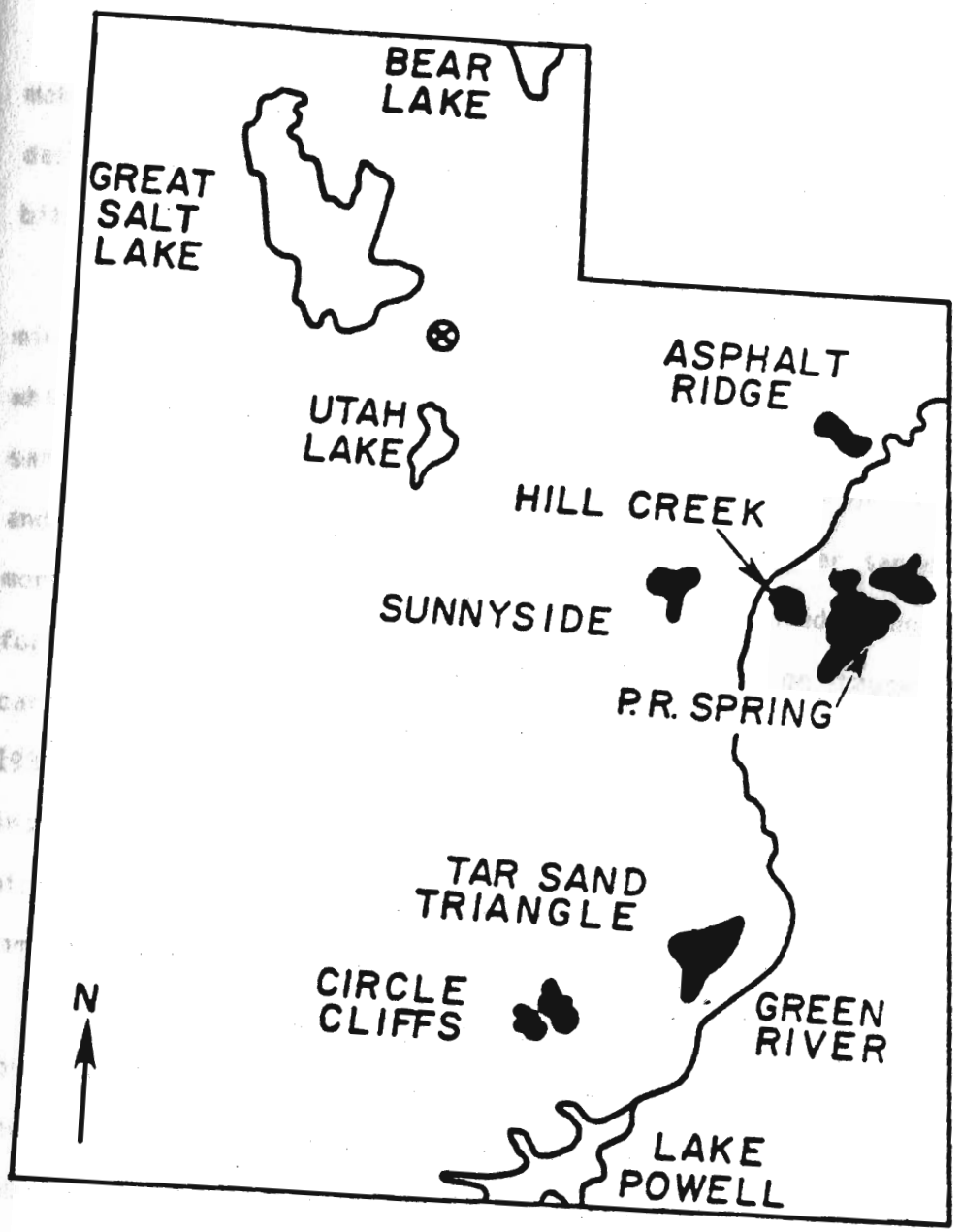


Figure 1. Location of the major Utah tar sand deposits

Tar Sand Properties

A knowledge of tar sand characteristics is essential for decision-making concerning hydrocarbon recovery. In this section, some of the data, relating to the physical properties of tar sand and its extracted bitumen, will be briefly reviewed.

Tar sand consists of an inorganic phase, usually sand or other mineral matter, and an organic, hydrocarbon-like phase, or bitumen, which either partially or completely envelops the sand particles. The sand is predominately α-quartz. However, small amounts of clay, calcite, and dolomite have been observed in some of the Utah tar sands. Furthermore, the sand tends to exist in either a consolidated or unconsolidated form, depending on the extent of cementing action caused by clays, carbonates, or other minerals present in the tar sand (Cortes, et al., 1982). The particle size distribution of Utah tar sands tends to indicate a smaller percentage of fines, classified as less than 40 microns, than the fines percentage of Canadian Athabasca tar sands of a similar class (Camp, 1974).

One of the major contrasts between Utah tar sand and the more abundant Athabasca tar sands of Alberta is the connate water content. The Canadian tar sands are believed to have a small water film separating the bitumen phase from the inorganic phase, Moreover, this water film is theorized to aid the hot-water, bitumen-extraction process (Camp, 1974). On the other hand, Utah tar sands are considered to contain negligible amounts of connate water, making them less amenable to the proven bitumen separation processes for Canadian tar sands (Sepulveda, et. al., 1978). This lack of connate water does not seem to

affect the ability to process the tar sand using the University of Utah's thermal process (Jayakar, 1979).

Bitumen is the term that refers to the highly viscous hydrocarbon substance which surrounds the inorganic phase of tar sand. The average bitumen content of tar sands tends to vary greatly from one deposit to another, and even within the same deposit. Utah bitumens tend to be more viscous than the Athabasca bitumen (Misra et al., 1981), but the Uintah Basin tar sands also contain substantially less sulfur (Bunger, et al., 1979). Table 4 shows some of the physical and chemical properties of the bitumen from selected Utah tar sands and from Athabasca tar sands. Figure 2 illustrates the viscosities of various tar sands as a function of temperature (Misra, et al., 1981).

Several conclusions can be surmised from the available data concerning physical-chemical properties of Utah tar sands. First, the relatively small connate-water content of Utah tar sands suggests that the conventional hydrocarbon processes used for the Athabasca reserves will need to be altered or replaced in order to successfully process most of the Utah reserves. Second, the bitumen, though highly viscous when compared to most conventional crude oils, is capable of being upgraded to a suitable crude for refinery feedstock replacement. Third, because of the lean nature of the tar sands, ranging between 1 and 17 weight percent (Wood et al., 1972), attention must be given to the large amount of inert sand that will need to go through the process. And fourth, though not previously mentioned, Utah tar sands present significant obstacles, with respect to geography, terrain, and water, which must be taken into account before finding a suitable process. These concerns are discussed in detail by Daniels, et al. (1982).

Table 4*

Physical-Chemical Properties of
Extracted Tar Sand Bitumens

Bitumen Source	N W Asphalt Ridge	P.R. Spring	Sunnyside	Tar Sand Triangle	Athabasca
Average Molecular Weight (VPO in benzene)	668	820	854	578	568
Specific Gravity (60/60)	.970	.998	1.003	.992	.989
API Gravity (°API)	14.4	10.3	9.5	11.1	11.6
Viscosity (20°C) kilopoise	30	325	NA	13.	6.4
Carbon (wt%)	85.2	84.4	86.3	84.0	82.5
Hydrogen (wt%)	11.7	11.0	11.1	10.1	10.2
Nitrogen (wt%)	1.02	1.00	0.80	0.46	0.47
Oxygen (wt%)	1.1	2.2	1.4	1.1	1.7
Sulfur (wt%)	0.59	0.75	0.35	4.38	4.86
C/H atomic ratio	0.611	0.641	.654	0.696	0.674
Carbon residue(wt%) (Ramsbotton)	3.5	12.5	14.1	21.6	16.11
Asphaltenes (wt%)	6.3	16.0	20.6	26.0	16.4

*Bunger, et. al., 1979

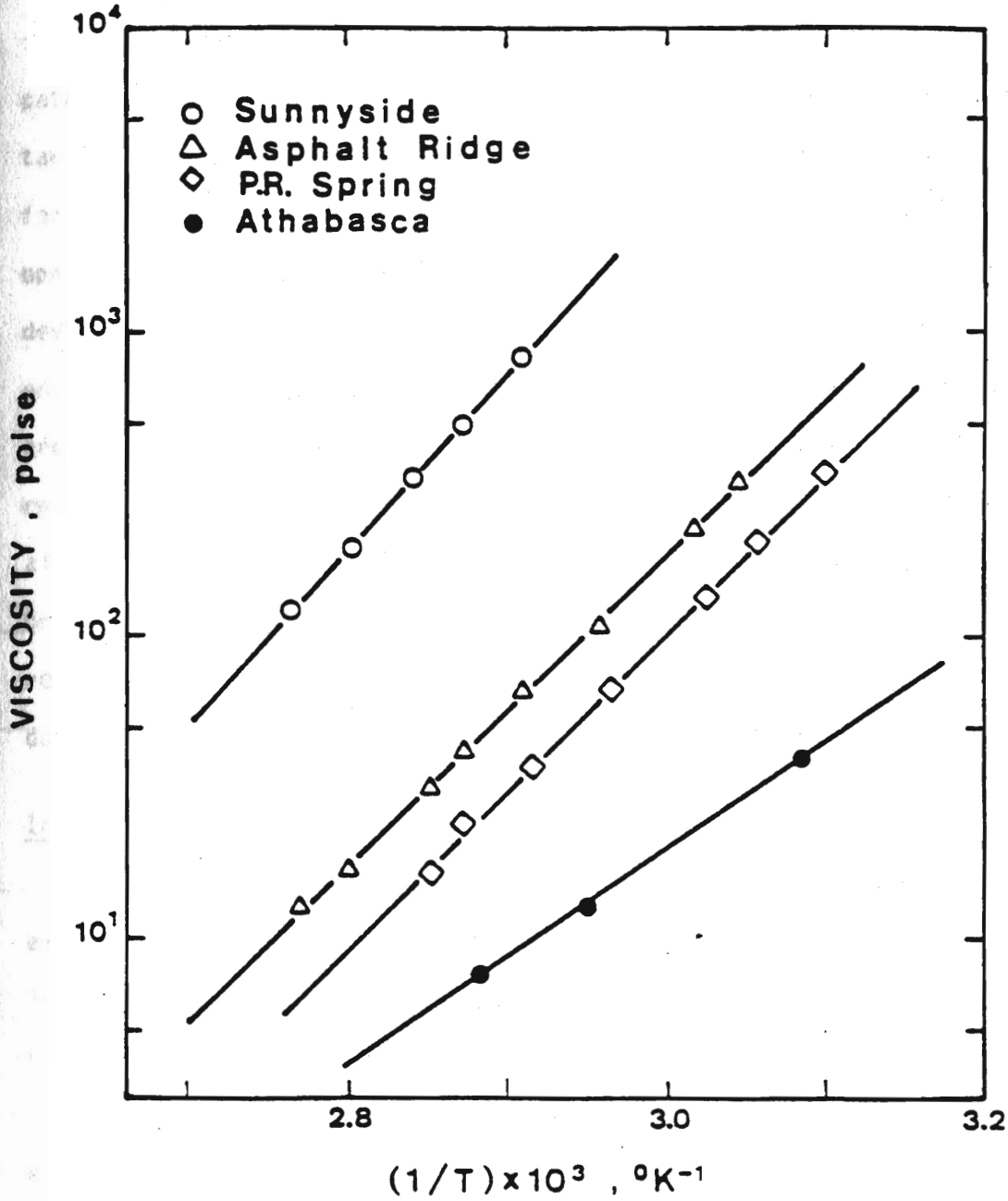


Figure 2. Effect of temperature on the viscosity of various bitumens

Methods for Tar Sand Processing

Most hydrocarbon-recovery processes for tar sands can be categorized as either in-situ processes, or as processes involving mined tar sands. Canada has been the site of several commercial operations for tar sand processing, consisting mostly of surface-recovery operations. However, activity has recently heightened towards the development of processes which will be both economically and environmentally feasible for the recovery of hydrocarbons from foreign and domestic tar sands. As of 1982, there were at least sixty tar sand related pilot plant studies being conducted in the U.S. alone (Meyer, et al., 1982). Some of the projects involving Utah tar sands will be briefly reviewed, but an exhaustive survey will not be attempted. The reader is referred to the thesis by Wang (1983) for a more detailed description of the different tar sand recovery techniques.

In-Situ Recovery

The in-situ recovery techniques that have received the most emphasis, relating to possible application to Utah tar sand reserves, include steam flooding, (Dorrence et al., 1981), and in-situ combustion, (Johnson et al., 1978).

In-situ recovery methods must be applied to tar sand deposits whose excessive overburdens prevent the use of surface recovery operations. Much of the technology developed for secondary and tertiary recovery of conventional crude oil is now being applied to tar sand deposits. The intent of these methods is to decrease the viscosity of the bitumen to allow its movement to production wells. Significant factors influencing the success of in-situ processes include the permeability of the

deposit, the nature of the overburden, the magnitude of the viscosity of the bitumen, and the chief characteristics of the reservoir fracture system.

Because the relative permeability of the reservoir approaches zero when the saturation of the bitumen exceeds about 75 percent, the injection of viscosity-reducing fluids, such as steam, hot water, hot gases, air or solvents, poses a serious problem which will probably require an artificial enhancement of the permeability of the deposit. (Marchant et al., 1979). An alternative approach would be to heat the bitumen by other methods, such as by induction or microwave heating.

Surface Recovery

The different surface-extraction processes for the recovery of oil from tar sand deposits can be categorized into four general areas. These are solvent processes, hot water and solvent-assisted water processes, thermal processes, and other, nonrelated processes. These categories will be briefly reviewed with most of the emphasis placed on thermal methods.

Not until recently has there been an emergence of a large variety of proprietary processes being studied for large-scale surface extraction of tar sands. Selection of the best surface recovery process will depend on variables such as specific characteristics of the tar sand deposit, type of solids handling required, degree of pretreatment needed, physical characteristics of the tar sand, capital and operating costs, and other technical and economic factors. A review of the different categories of surface processes is given by Irminger et al. (1983).

Solvent-assisted water processes consist of several general steps. First, the tar sand is crushed to a feed of minus 1/4-inch or less. Then the tar sand is conditioned in a water, alkali-assisted step to physically and chemically disengage the bitumen from the tar sand. An organic solvent is then added to collect the bitumen. A flotation step follows to separate the bitumen from the alkaline water and sand. The sand and water slurry tailings from the flotation step are then processed to produce water and a damp sand. The tar sand concentrate is then processed to recover the bitumen.

The above method resembles the Canadian Clark Hot Water Process which has been used successfully on the tar sands of Alberta, Canada. The main disadvantages of the solvent-assisted water process, applied to Utah tar sands, are large water and solvent requirements, and the fact that Utah tar sands tend to be oil-wetted, which is the opposite of the water-wetted tar sands of Canada which have proven to be quite amenable to the water process.

A modified hot water process was operated on a pilot scale by Enercor and the State of Utah in order to demonstrate the functionality of this method on a continuous basis using a variety of different Utah tar sands. The results of that work are discussed by Hatfield et al. (1982). Forster (1983) discusses the development of a 25 barrel per day test unit using a modified version of the process to recover oil from New Mexico tar sands.

Solvent extraction processes include use of solvents either lighter or heavier than water, and can be summarized in a few general steps. The bitumen concentrate or sized tar sand is mixed with a suitable solvent,

usually a fluoro-chlorinated organic for heavy solvent processes; or hexanes, heptanes, kerosene, toluene, and others, for light solvent processes. The sand, bitumen, and solvent micelle is then passed through a series of countercurrent extraction stages. Various operations are employed to remove and dry the coarse and fine sand, and to recycle the solvent. Fines, especially those eminating from clays, can lead to serious emulsion problems. Following the removal of the fines from the concentrate, the bitumen is then sent to the upgrading process. A major difficulty with a solvent process is the selection, cost, and loss of the solvent.

Thermal Processes

The generalized operation of a thermal process for bitumen recovery from tar sands entails mining and sizing of the raw tar sand, the thermal process itself to remove the bitumen from the sand, use of supporting processes for product separation and fractionators, flue gas treatment, and heat recovery. The key factor determining the degree of success of a thermal process is the optimization of the thermal efficiency.

Presently, several processes are being investigated as to their application to U.S. or Utah tar sands. Some processes available for license include the Tosco (rotary kiln), ERCO (fluid bed), and the screw-type Lurgi process (Weiss, 1983).

There are several significant features of a thermal process which separate it from the other surface extraction processes. First, the thermal process supplies the opportunity for power cogeneration by the use of waste heat. Second, the thermal process is a partial upgrading

step that will replace, to a degree, the delayed coking steps of the other processes. And third, water requirements for most thermal processes are modest compared to those of the water or solvent-assisted water processes.

Jayakar (1979), Venkatesan (1980), and Wang (1983) give adequate reviews of thermal processes investigated for the above-ground oil recovery from tar sands. This review will briefly cover the development of the University of Utah thermal process.

The University of Utah Thermal Process

Weeks (1977) presented the first work relating to the original process, which consisted of a two-staged fluidized-bed reactor. Coked sand from the upper bed was transferred by gravity to the lower, combustion bed, while the oil products of the coking bed passed through a product recovery system. Sensible heat of the hot gases, leaving the combustion bed and fluidizing the upper bed, was the medium used to provide the major portion of the energy required for bitumen pyrolysis in the upper bed. Propane, or other combustible gases, were used to supply any extra energy required for pyrolysis. Runs were made using tar sand from the Tar Sand Triangle deposit of Utah.

The process was then modified by Jayakar (1979) to permit more efficient thermal coupling of the two fluidized beds, by employing the technology of liquid-metal heat pipes. A single liquid-potassium heat pipe was installed to extract the heat from the combustion bed needed to maintain the necessary temperature for pyrolysis in the upper bed. Data on product yields and characteristics were obtained for tar sands from the Tar Sand Triangle and Asphalt Ridge deposits of Utah. Processing

studies using Sunnyside tar sand were also made on the same system by Hanks (1979).

Bezama (1983) modified the process further by separating the fluidizing gas flows for the two beds. A digital data acquisition system was installed to permit better study of the dynamic behavior of the process.

A digital control system was then installed and tested by Hsu (1983) to provide better control of both the pyrolysis-bed temperature and the levels of the fluidized beds.

The latest modifications of the process completed in this research, entail the design, construction and operation of a larger unit using three liquid-metal heat pipes. Moreover, the product recovery system was modified to provide more efficient synthetic-crude collection. Addition of an air preheater and an additional fluidzed-bed heat-recovery section was also completed. The previously developed data acquisition and control systems were then adapted to the scaled-up reactor. Runs were made with four different Utah tar sands to demonstrate the controllability and efficiency of the scaled-up process.

CHAPTER III

PROCESS AND APPARATUS

Process Description

Figure 3 shows the basic material flow diagram for the process used in the experimentation. Tar sand is fed at a constant rate into the pyrolysis bed, where the bitumen is converted into vaporized pyrolysis products and coked sand. The gaseous phase, consisting of the fluidizing gas and the vapor products, leaves the pyrolysis section and passes through the product recovery section, where liquid crude is recovered. Meanwhile, the coked sand leaves the pyrolysis bed and enters the combustion bed, where it reacts with fluidizing air to combust the coke layer. The gaseous combustion products and unreacted air are vented out of the combustion section, while the spent sand leaves the combustion section and enters the heat recovery section. A fluidizing air stream enters and leaves the heat recovery section unreacted, and the sand exits this section in a cooled, spent state.

A more detailed diagram of the University of Utah thermal process is shown in Fig. 4. Ground tar sand, usually between number 9 and 24 Tyler mesh, is fed by a screw conveyor which is driven by a continuously variable D.C. motor. The tar sand falls from the top of the feeding column into the upper fluidized bed. A nitrogen purge is introduced above the feed column to prevent oil vapors from plugging the feeding system. The coking bed, which is fluidized with a steady flow of

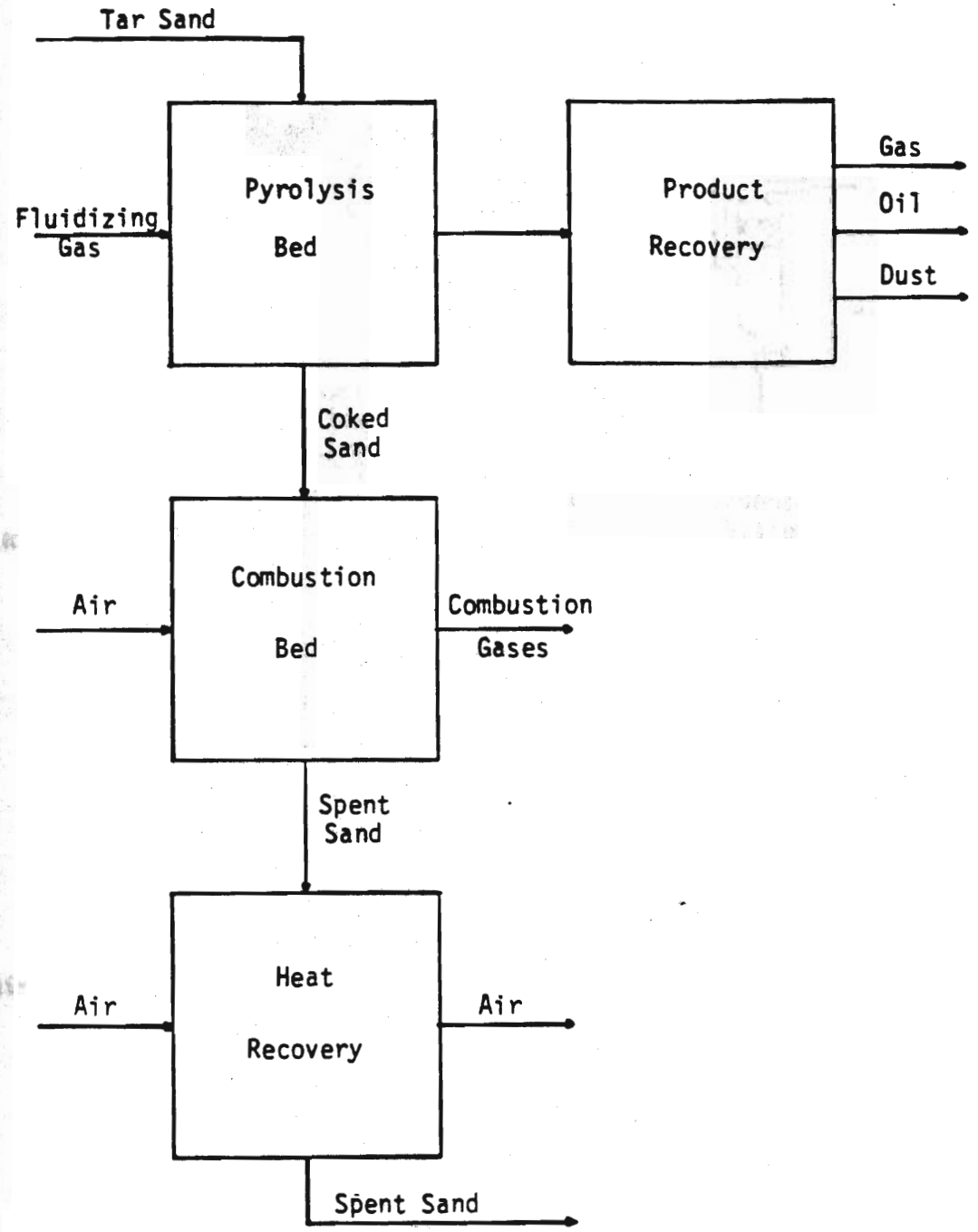


Figure 3. Basic material flow for the University of Utah thermal process

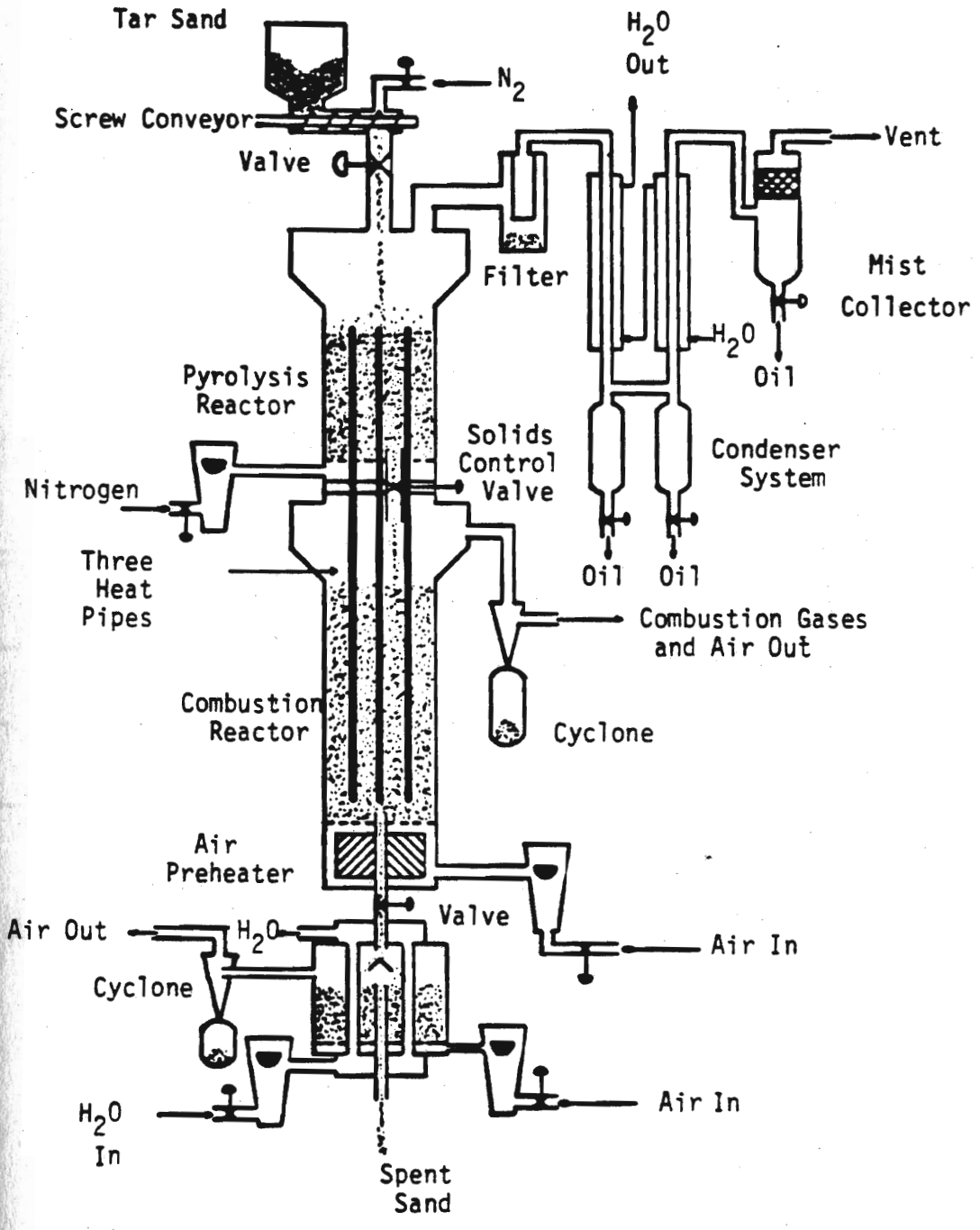


Figure 4. The University of Utah thermal process

nitrogen, is maintained at a temperature high enough to pyrolyze the bitumen layer on the tar sand. The fluidizing and purge nitrogen, oil vapors, and gases then leave the coking bed and pass through a 40-micron, cylindrical stainless-steel filter to remove most of the entrained fines. The vapors and gases then pass through two water-jacketed double-pipe condensers, where a portion of the vapors condense on the walls and flow down to collecting vessels at the base of the condensers. The uncondensed vapors flow on through a mist collector, which consists of a reverse-flow depth-type fibermist eliminator and a base to collect the coalesced liquid phase. The gases and any remaining uncondensed vapors are then occasionally sampled and vented.

The coked sand formed in the pyrolysis bed is transferred down to the fluidized combustion bed by gravity via a solids control valve. This bed is maintained at a temperature to permit spontaneous combustion of the coke layer with the fluidizing air. Three heat pipes, with liquid potassium as the working fluid, are used to thermally couple the upper pyrolysis bed to the lower combustion bed. The exothermic reaction in the lower bed produces gaseous combustion products and heat; the former exit with the unreacted fluidizing air through a cyclone and vent, while much of the latter is transferred up to the pyrolysis bed at a sufficient rate to maintain the required upper-bed temperature. The remainder of the heat of combustion remains with the spent sand as sensible heat.

To remove this remaining heat from the spent sand, the hot sand is sent, again by gravity, from the combustion bed through an air pre-heating section, a lower solids flow control valve, and into a third fluidized bed. This bed has four vertical copper cooling tubes, which

are supplied with a regulated flow of water to pick up the required heat. The cool spent sand exits the third bed, which is fluidized with air, through a simple stand pipe that controls the level of the bed.

Figure 5 shows the actual physical arrangement of the tar sand processing unit drawn to scale to give an indication of the relative sizes of the different sections. The entire unit, which is supported by an angle iron frame, has a height of fifteen feet and occupies parts of two different floors of the Department of Chemical Engineering.

Primary Reactor Design

Fluidized Bed Reactors

The original tar sand processing unit was designed to process tar sand at a rate of about 5 lb/hr, and since one of the objectives of this project was to have a solids processing capacity of at least 15 pounds per hour, the inside diameters of the two main reactor sections were scaled up from 2.157 inches to 4.260 inches. Assuming similar fluidized bed heights, bed densities, and mean residence times, this would allow for a tar sand processing capacity of nearly 40 lbs/hr. An inside diameter of 4.260 inches was chosen because of material availability and the fact that it could easily handle the design sand flow of 15 lbs/hr.

The actual constructions of the combustion reactor and the pyrolysis reactor are shown in Figs. 6 and 7, respectively. The shells were made from 4-inch pipe sections (type-304, schedule-10S, stainless-steel) arc-welded to concentric weld reducers (8 to 4-inch nominal, type-304, schedule-40, stainless-steel) which were welded to lengths of 8-inch pipe of similar material (schedule 10S). This resulted in a

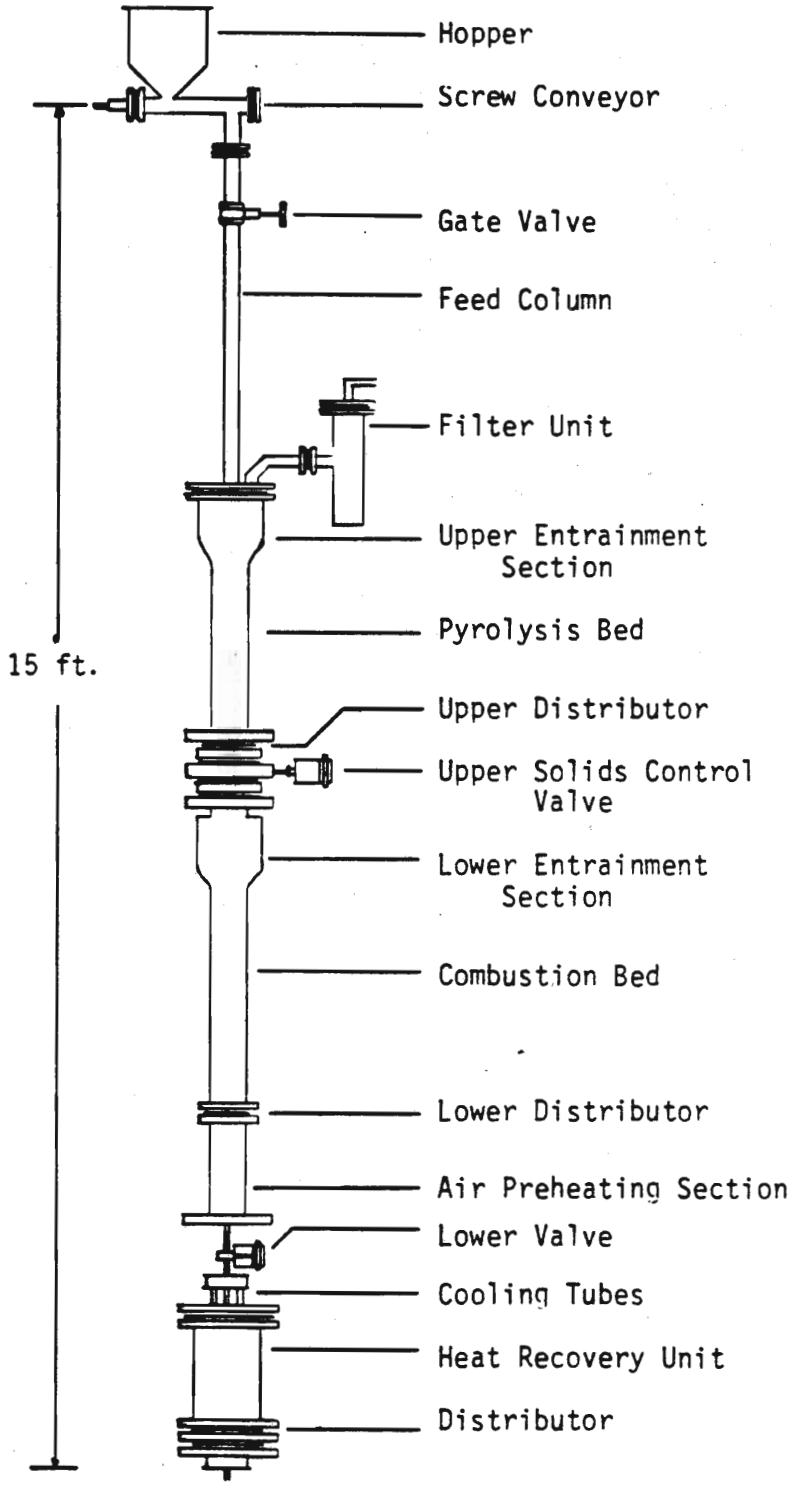


Figure 5. The assembled reactor

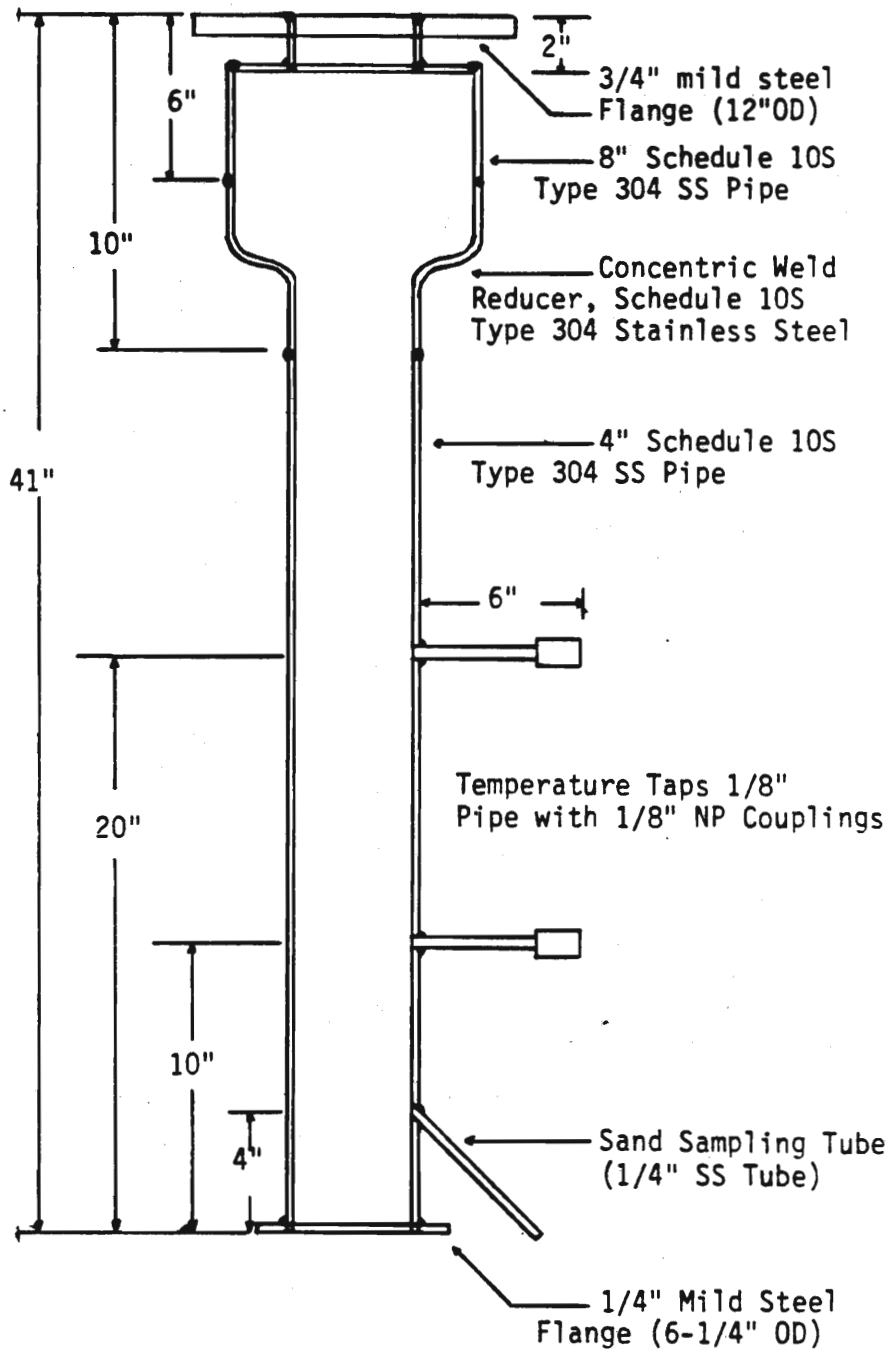


Figure 6. The combustion reactor

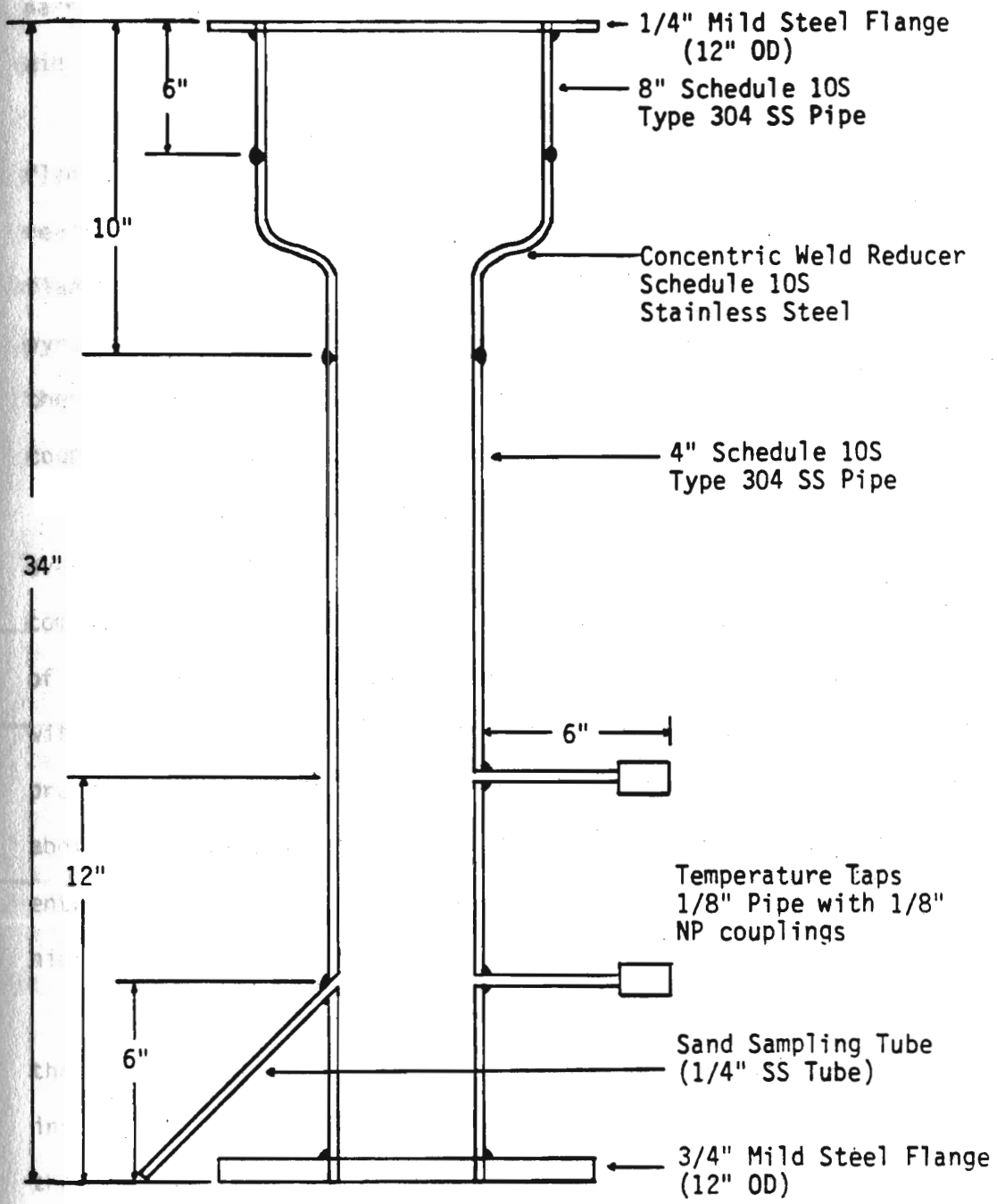


Figure 7. The pyrolysis reactor

narrow section to contain the fluidized bed and an expanded section to aid in the disengagement of entrained fines.

The bottom of the combustion reactor was fitted with a mild steel flange for attachment to the air preheating section. The top of this reactor was reduced back to a 4-inch nominal size and fitted with a flange fabricated from 3/4-inch mild steel for attachment to the pyrolysis reactor. Also welded to the combustion reactor were two thermocouple ports, fitted to hold 1/16-inch stainless-steel thermocouple probes, and a solids-sampling port fitted with a ball valve.

Both reactors were wrapped with several layers of silica cloth, over which lengths of 16-gauge nichrome heating wire were wound. The combustion bed had two layers of heating wire installed in anticipation of larger heating requirements; whereas, the pyrolysis reactor was wound with a single layer of nichrome wire. Since experience with the previous processing unit indicated that the product vapors needed to be kept above 400°C in order to prevent plugging of the product exit line, the entrainment section of the pyrolysis reactor was wound with 22-gauge nichrome heating wire.

The entrainment and reactor walls of the pyrolysis section were then wrapped with 1.5-inch and 3.5-inch thick layers of ceramic fiber insulation, respectively. This insulation was chosen because of its low thermal conductivity and its resistance to high temperatures. The fiber insulation layer was covered with a 1.5-inch thick layer of calcium-silicate-block pipe insulation and secured with an aluminum moisture-resistant shell for support. The combustion section was insulated in a similar manner and ceramic tubes were used to encase the heating wire as it entered and exited through the layers of the insulation.

Before the two main reactor sections were installed, heating tests were conducted to analyze the durability and performance of the completed sections. This was accomplished by connecting the heating elements to a power source and allowing the sections to reach a specified steady-state temperature. Calculations of heat losses were made using measured interface temperature, known thermal conductivities, and the physical dimensions. The results of these tests indicated that the two sections would be able to reach the temperatures required for the thermal process.

Distributors

The gas distributors function to evenly distribute the fluidizing gases and to support the sand bed. Figures 8 and 9 show the design of the distributors for the pyrolysis bed and the combustion bed, respectively. These were made by sandwiching a layer of 200 Tyler mesh stainless-steel screen between two disks fabricated from 1/16-inch stainless-steel plate. The discs were arc-welded along the edges to secure the screen and prevent gas leakage. Figure 8 also illustrates how the heat pipes were positioned to symmetrically compartmentalize the fluidized bed.

Preliminary tests of the distributors indicated that fluidization at design bed hold-ups and air flow rates was highly nonuniform due primarily to a negligible pressure drop across the screen. A general rule of thumb in distributor design, as discussed by Kunii and Levenspiel (1969), states that the pressure drop across the distributor is roughly 10 percent of the pressure drop across the bed, with a minimum in all cases of about 35 cm water. It was then decided to

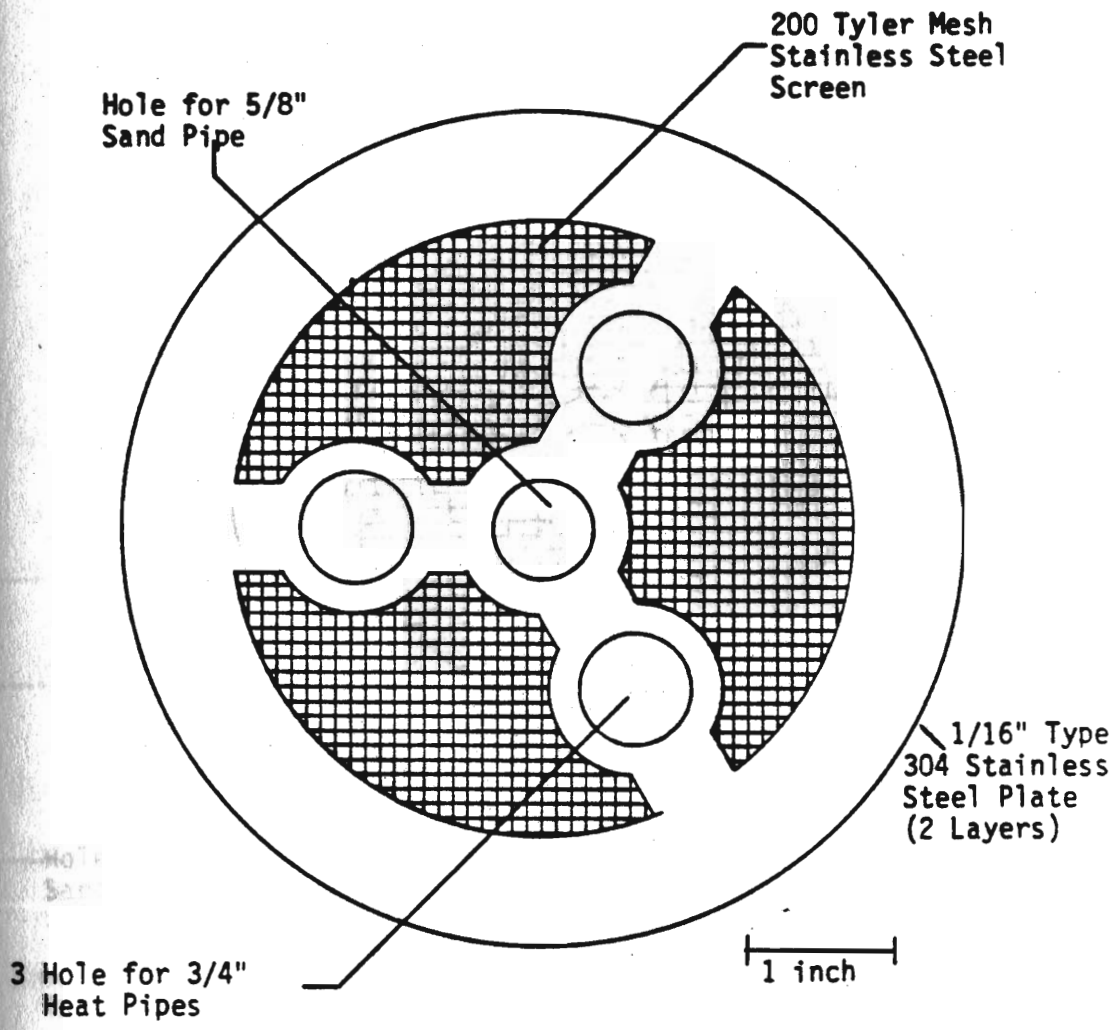


Figure 8. The distributor for the pyrolysis bed

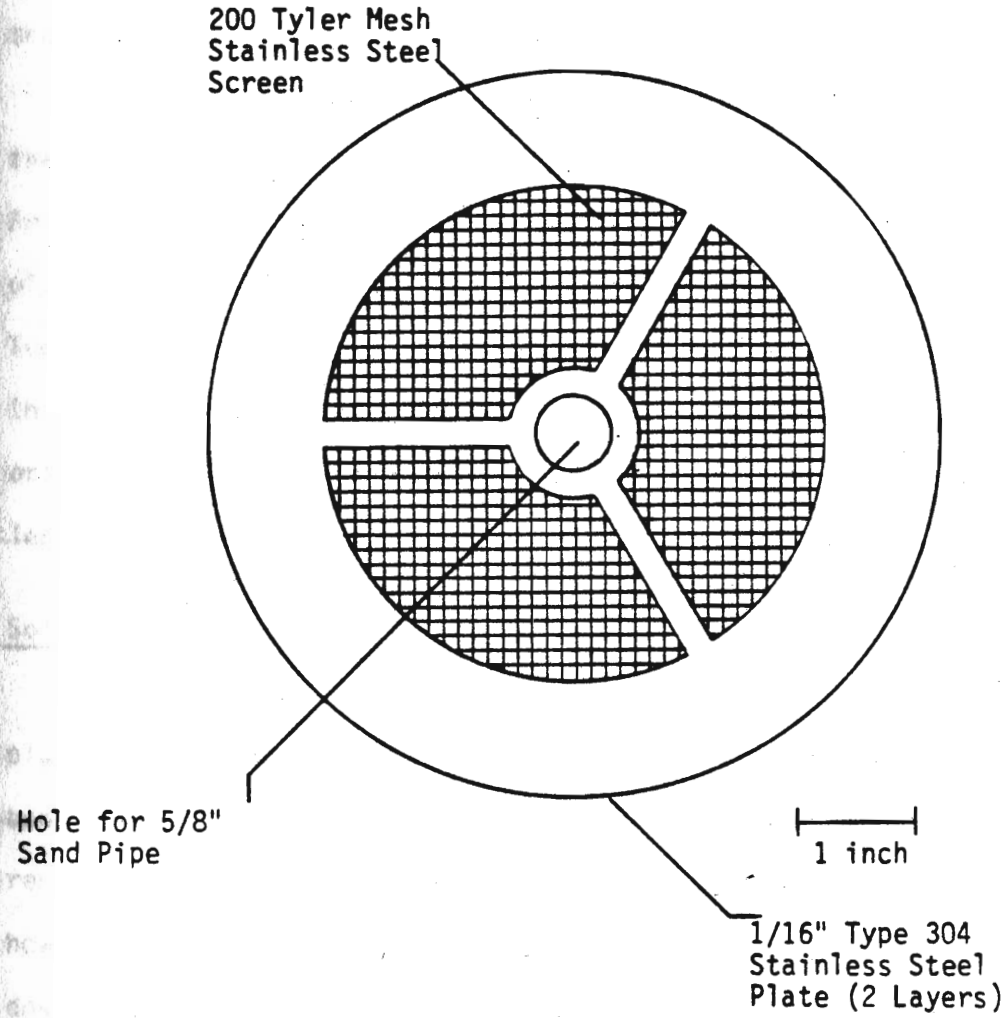


Figure 9. The distributor for the combustion bed

construct perforated plate distributors to be placed directly below the steel screen distributors which would allow for a greater pressure drop, while the screen would prevent the flow of sand back through the orifices of the plate.

The design of the perforated plates was accomplished using orifice theory and the procedure recommended by Kunii and Levenspiel (1969). From the calculations made, it was determined to make the perforated plates from 1/16-inch stainless-steel plate, with one orifice per cm^2. The inside diameter of the orifices were calculated to be 1 mm. Installation of these perforated-plate distributors directly under the original wire-mesh distributors resulted in more ideal fluidization with less channeling.

Solids Flow

The original tar sand processing unit experienced substantial plugging problems relating to the transfer of solids from the fluidized beds, where plugging of the 3/8-inch OD downcomers was a common occurrence. The new unit uses essentially the same solids transfer system; however, the 3/8-inch OD downcomers have been replaced with 5/8-inch OD downcomers, representing a four-fold increase in cross-sectional sand flow area. This increase in sand flow area resulted in uninterrupted solids flow from both fluidized beds.

The solids flow through the stainless-steel downcomers, which extend 2-inches up from the bottom of the beds and have an ID of 1/2 inch, and are regulated by slide valves operated by Honeywell pneumatic actuators. A complete description of the actuators and their operation is given by Weeks (1977). Design of the upper solids control valve for

bed level control of the pyrolysis section is shown in Fig. 10. This figure shows the section of the process which is termed the divider plate. The divider plate's three functions are first, to support the three heat pipes; second, to separate upper and lower fluidizing gas flow; and third, to house the upper solids control valve. A 1/2-inch type 304 stainless-steel rod operates through a 17/32-inch hole that was drilled midway through the side of the plate. The valve stem slides past the sand pipe hole to completely shut off any sand flow; moreover, a 3/8-inch hole was drilled behind the sand pipe and through the bottom half of the plate to prevent any build-up of sand which would jam the valve. A 1/4-inch hole was also drilled through the side of the plate into the area where the slide valve shuts off the flow. This port was made in order to provide a purge line to inject nitrogen for the dislodgement of any minor plugs. Two 3/8-inch actuator support rods were threaded into the side of the divider plate for support of the valve actuator. Other holes drilled into the divider plate include 3/8-inch holes which house the bolts that connect the two main reactor sections, and three holes which tightly house the 3/4-inch heat pipes.

Heat Pipes

The heat required in the pyrolysis bed to heat the incoming tar sand to the required pyrolysis temperature, to crack the bitumen, and to heat the incoming fluidizing gas is provided entirely by three heat pipes which transfer heat from the lower combustion bed. Runs made by Jayakar (1979) on the previous unit, which used only one heat pipe and was designed to process 5 pounds of tar sand per hour, indicated that

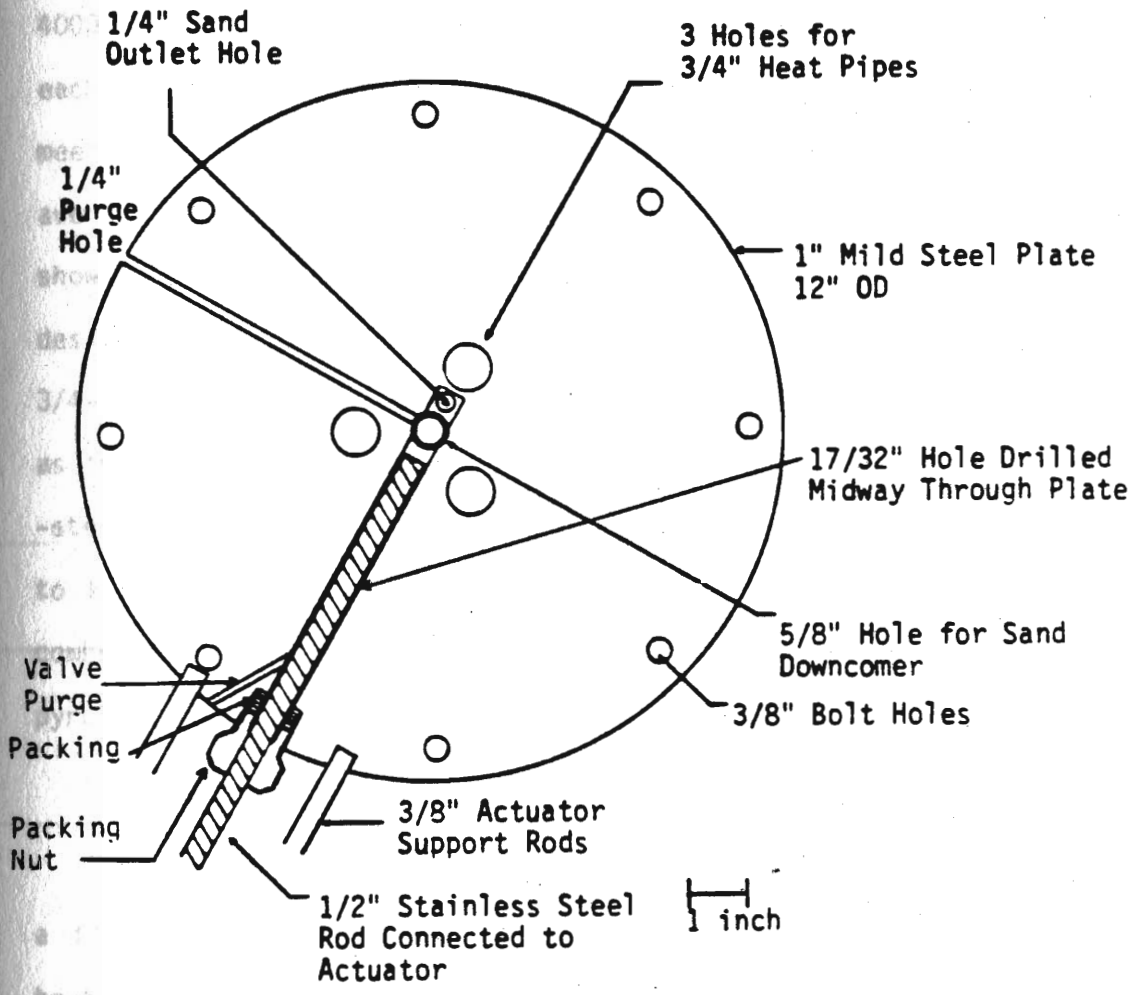


Figure 10. Design of the upper solids-control valve

the maximum heat transferring capacity of 3695 Btu/hr at 500°C adequately provided the necessary heat transfer. Taking this into consideration and noting that the new design was for 15 pounds per hour, 4000 Btu/hr was determined to be the design-heat-transfer capacity of each of the three new heat pipes. The heat pipes were then designed to meet specifications of overall heat-transfer load for the process and average temperatures of the condensing and evaporating sections. Table 5 shows the specifications which were given to Thermocore, Inc., for the design of the three heat pipes. Thermocore built the heat pipes out of 3/4-inch OD, type 304 stainless-steel tubing and chose liquid potassium as the most suitable working fluid. The three heat pipes have stainless -steel pinch-on clamps attached, which rest on top of the divider plate to keep the heat pipes approximately 1/2-inch above the bottom of the combustion bed and to give an exposed length of 18 inches into the pyrolysis bed.

Heat Recovery System

The heat recovery system consists essentially of two main sections, a fluidizing-air preheater for the combustion bed and a fluidized-bed heat exchanger to transfer the sensible heat of the hot-spent sand to a regulated flow of cooling water. Figure 11 shows the design of the air-preheating section, located directly below the combustion bed, which consists of a 5/8-inch OD stainless-steel tube with four 1/8-inch-thick copper fins encased in a 15-inch section of 4-inch nominal, type 304, stainless-steel, schedule 10S pipe. The base of the preheater is fabricated from a 1/2-inch-thick, mild-steel plate, and functions as a

Table 5

Heat Pipe Design Specifications

Variable	Value
Position	Vertical
Length	61 inches
Outside Diameter	0.75 inch
Condensing Section Height	18 inches
Evaporating Section Height	28 weeks
Adiabatic Section Height	15 inches
Material	Stainless Steel
Temperature Limits:	
Evaporating Section	550 - 650°C
Condensing Section	450 - 550°C
Approximate ΔT	100°C
Power Requirement	4000 Btu/hr.

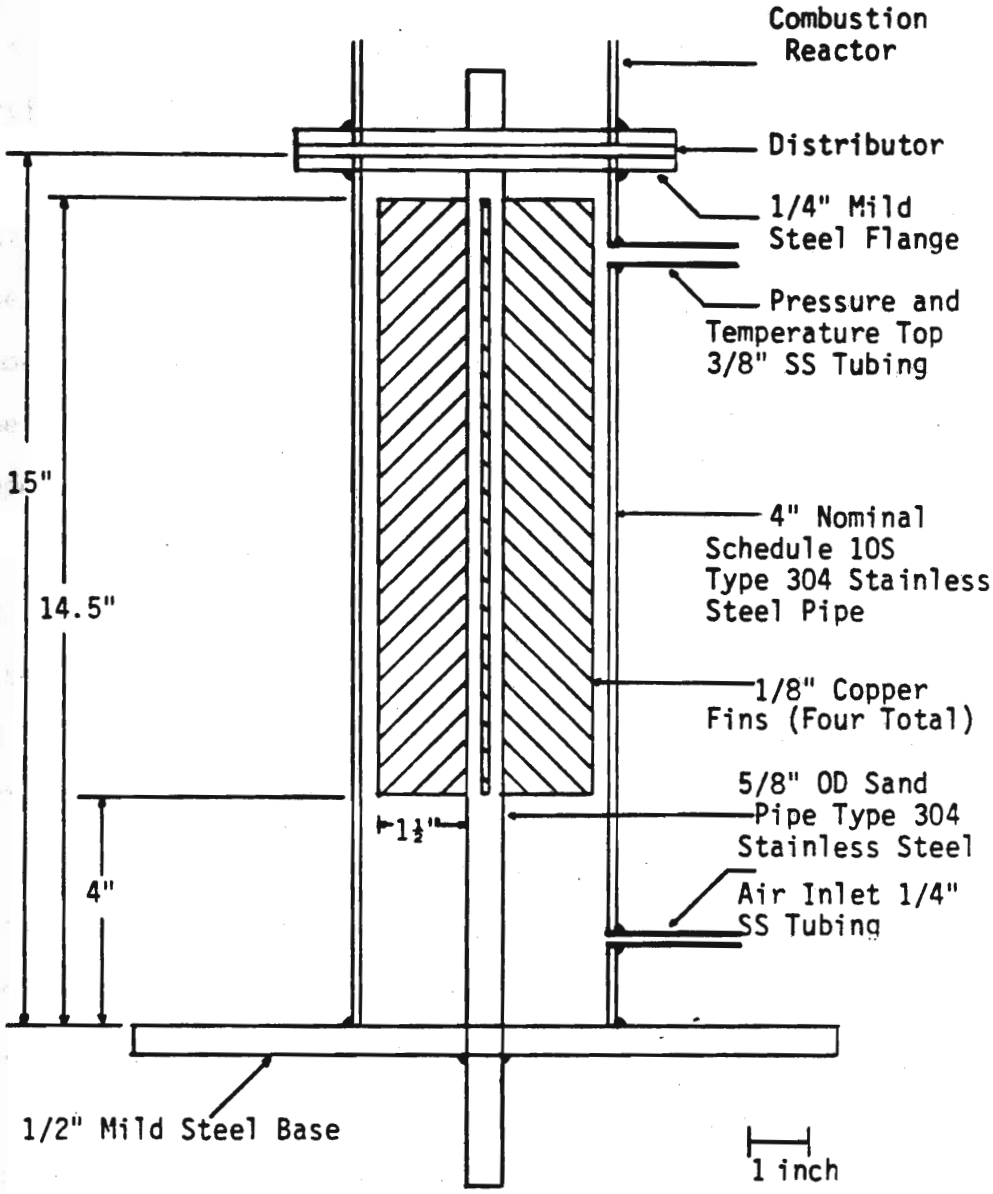


Figure 11. The air-preheating section

support for the tar-sand unit. A 1/4-inch, mild-steel flange is welded to the top of the preheater and attaches to the base of the combustion section to hold the distributor in place.

Fluidizing air enters the preheater via a 1/4-inch stainless-steel tube that is welded to the wall at a height of 1.5 inches from the base. A thermocouple port is also attached to this tube to allow for temperature measurement of the incoming air. A 3/8-inch stainless-steel tube was also attached at a height of 14 inches from the base for temperature and pressure measurement at that point.

The second section of the heat recovery system consists of a fluidized-bed heat exchanger with immersed, vertical cooling tubes. After specifying the fluidized-bed diameter; the number, size, material, and location of the cooling tubes; the sand flow rate; the inlet and outlet temperatures of the flowing sand and water; the characteristics of the sand; and the flow rate of the fluidizing air, the required height of the fluidized bed was calculated. The convective heat transfer coefficients were determined by the Hausen correlation (Perry, et al., 1973), and the Wender and Cooper correlation (Botteril, 1975). The design of the fluidized-bed heat exchanger is shown in Fig. 12. Control of the bed height was achieved by using a simple vertical overflow pipe; however, an air-tight vesssel had to be attached to the exit of the overflow pipe to prevent air from flowing out the overflow pipe instead of the cyclone. Ports made of 1/4-inch stainless-steel tubing and 1/8-inch NP couplings were attached to the heat exchanger to measure inlet and outlet water temperatures, fluidized-bed temperature, inlet and outlet air temperatures, and bed pressure drop. A distributor,

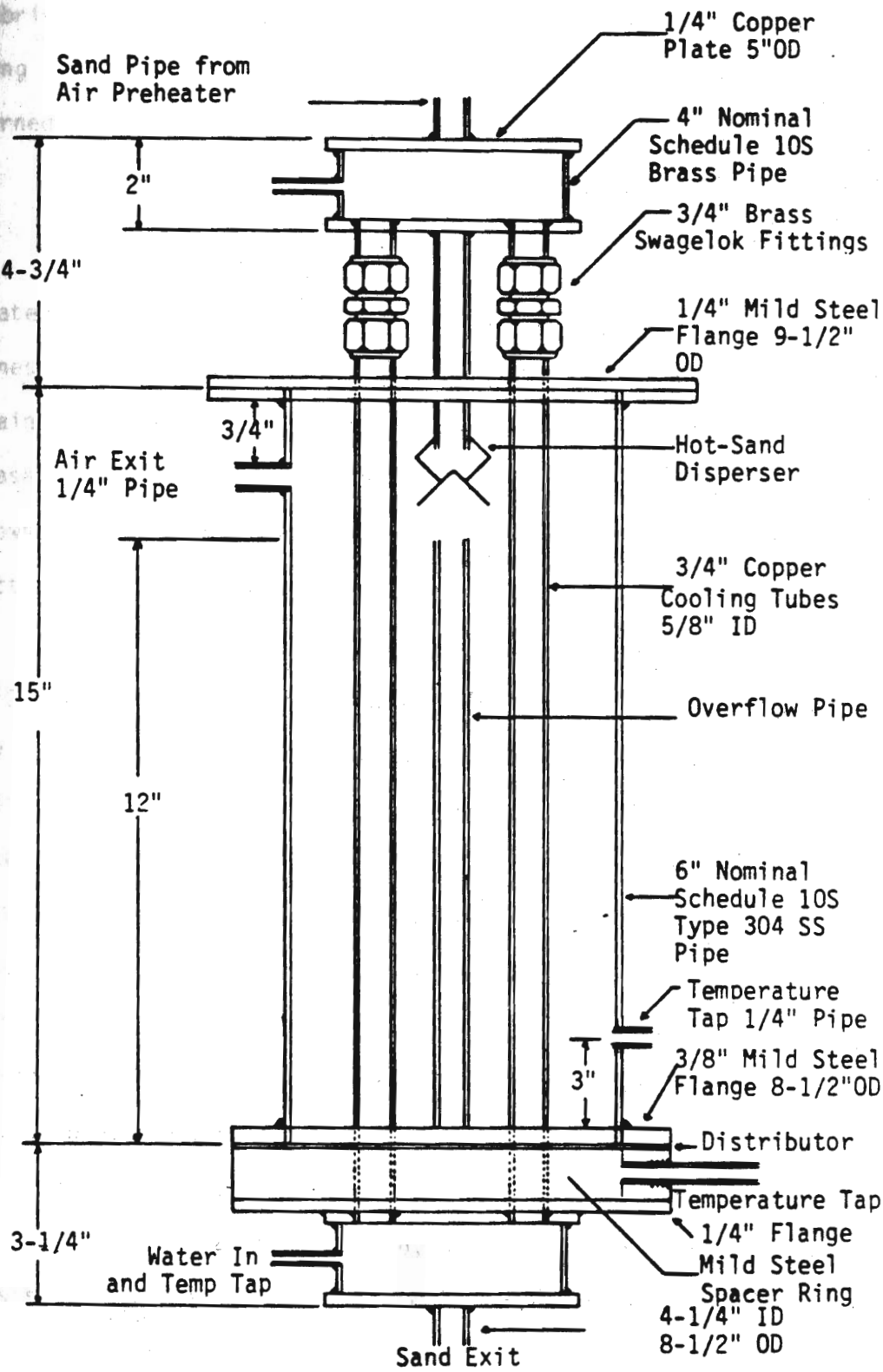


Figure 12. The fluidized-bed heat exchanger

fabricated in the same manner as the other distributors, and a spacer ring for even air-flow distribution, were secured between two flanges turned from 1/4-inch mild steel.

Solids Separation System

The solids separation system includes a heated exit line and a heated filtering section. Hot oil vapors, gases, and entrained sand fines flow through the exit line, then pass through a sintered-stainless-steel filter where the fines are separated from the gaseous phase. During a tar sand run, a filter cake of coked fines gradually grows on the surface of the filter until it falls and is retained in the section immediately below the filter.

Early runs with tar sand indicated that the exit line, which was originally fabricated from 1/2-inch-ID flexible-stainless-steel tubing, would eventually plug with collected coked-sand fines. To eliminate this problem the exit line was remade out of 2-inch-nominal pipe sections which were wrapped with silica cloth, wound with 16-gauge-nichrome heating wire, and insulated with a 1-inch layer of ceramic-fiber insulation. The exit line is welded to the top of the upper-entrainment section and connected by 1/4-inch flanges to the filter section. A port for measurement of the temperature inside the exit line is also provided for.

Figure 13 illustrates the configuration of the filter section which is connected to and supported by the exit line. The filter section consists of an outer shell made from 4-inch-nominal stainless-steel pipe that encases a sintered-stainless-steel filter element with a nominal pore size of 5 microns and a wall thickness of 1/16 inch. A 1/4-inch

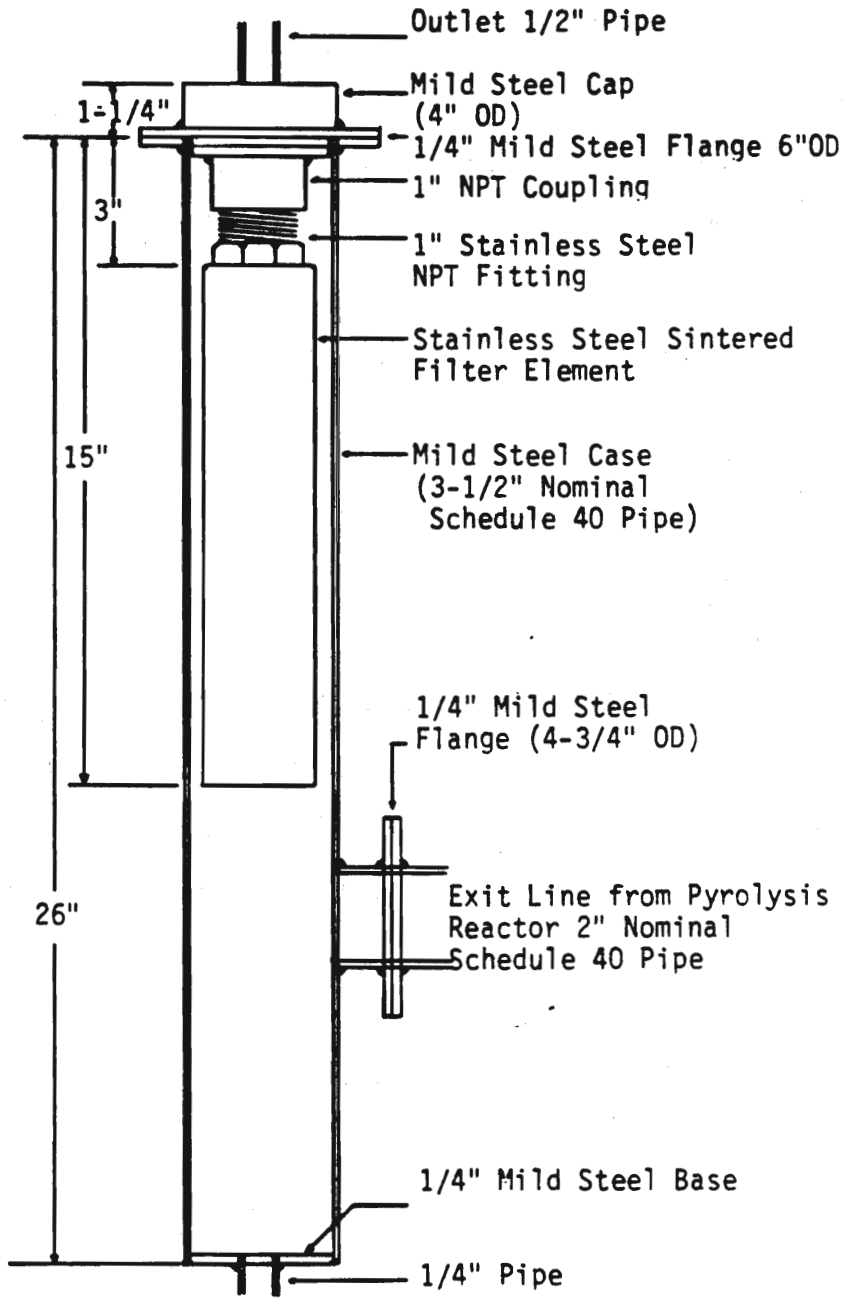


Figure 13. The filter section

gate valve was installed at the bottom of the encasement to allow for pressure relief and/or fines removal. The filter section is heated and insulated by a method similar to that used for the exit line. A thermocouple port is located at the top of the filter section for temperature measurement inside the metal filter in order to maintain the filter at a sufficient temperature to prevent condensation of oil vapors. The filter section is connected to the product recovery system by 1/2-inch nominal pipe fittings and Swagelok fittings.

Product Recovery System

The product recovery system consists of a series of two water-cooled double-pipe heat exchangers and mist-collection unit. The condensers were fabricated from copper tube sections with standard soldered fittings. The first exchanger has outer and inner tube sizes of 1-5/8 and 7/8 inches, respectively, and a length of 40 inches; whereas, the second condenser has pipe sizes of 1-inch and 1/2-inch, and a length of four feet. Water flows cocurrently with the gas flow for the two vertically-positioned condensers.

The mist collection unit consists of a reverse-flow depth-type liquid-gas coalescing element and a section to collect drained oil. Nitrogen, product gases, and entrained liquids in the form of a fine mist enter the base of the mist collector through a 1/4-inch-nominal pipe and are forced through the fibrous liquid extractor. The dispersed liquid phase collects on the random fibers. As the size of the droplets increases, the effect of gravity overcomes the force due to adhesion to the fiber surface and the force due to the upward velocity of the gas phase; consequently, the liquid falls from the extractor to the collection base.

The design and construction of the mist collector is illustrated in Fig. 14. Three sections of schedule-40S stainless-steel pipe with a nominal size of 4-inches were used to construct the mist collector. Flanges were made from 1/4-inch mild steel to connect the three sections. Cylindrical metal screens, with diameters equal to the inside diameter of the pipe sections, were welded to the top of the bottom section and to the bottom of the top section so that when the three sections are bolted together the fiber-mist-collecting element are held in place. One-fourth-inch NPT taps were placed in the pipe sections above and below the mist eliminator in order to measure the pressure drop across the random-fiber element. Gas flow enters and exits the mist collecting unit through 1/4-inch-nominal pipe sections threaded as shown. The same type of pipe is also threaded to the bottom of the unit and fitted with a 1/4-inch gate valve so the oil can be drained from the base.

Two types of materials are being used to provide the random-fiber material for the mist collector. Pyrex-brand filtering fiber worked very well but was also quite expensive to use. For this reason, layers of air-filter pads, made from spun-bonded polyester fiber, were used with layers of cellulose fiber to successfully knock out most of the mist.

Feeding System

The feeding system used in the scaled-up tar sand reactor is essentially the same as the system used in the previous research. A description of this system is given by Jayakar (1979). Although the scaled-up reactor is built to handle a larger feed rate, the reactor

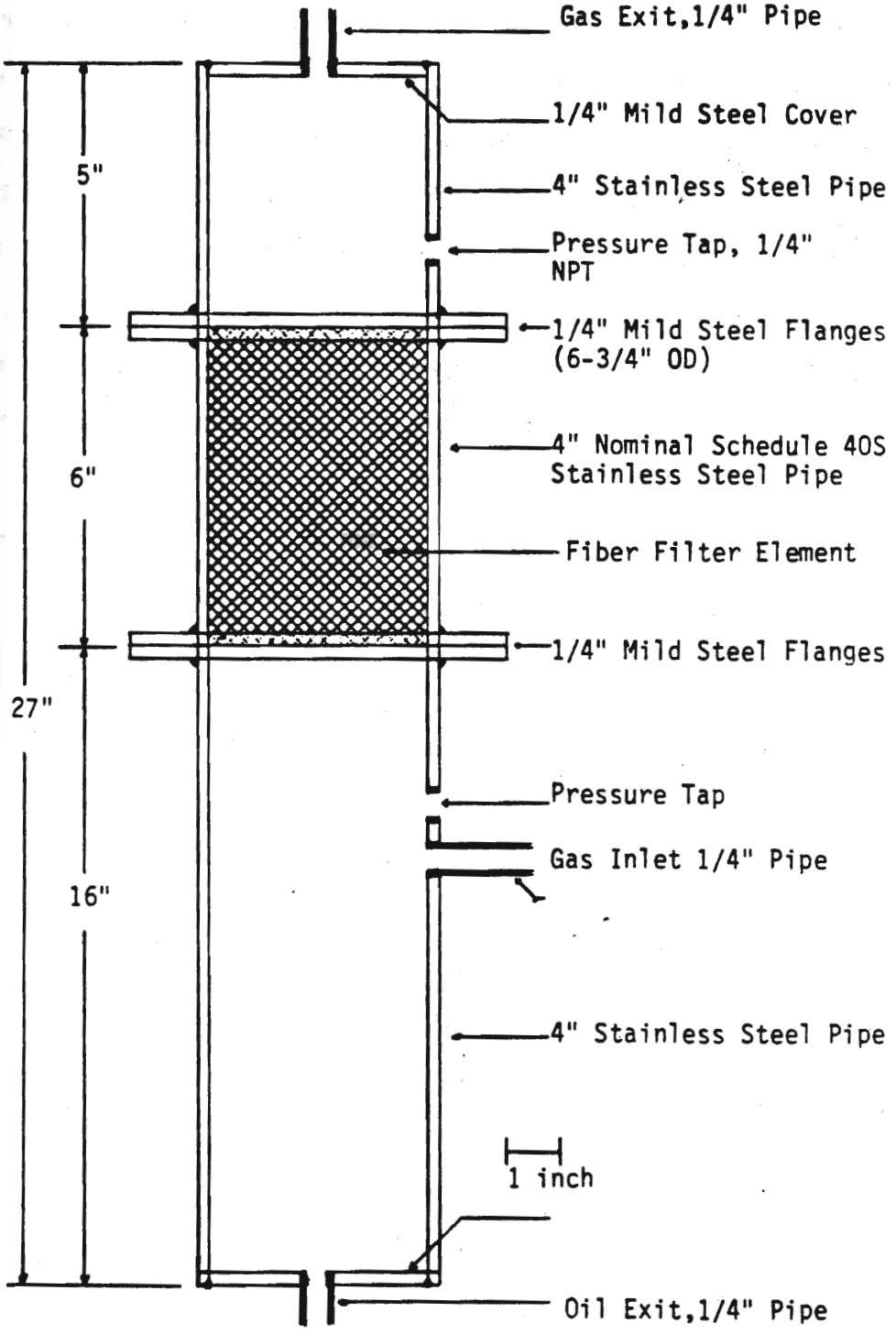


Figure 14. The mist elimination section

sections are considerably shorter than those of the previous unit because it was overdesigned with respect to height; hence, an extension of the feeding column, connecting the screw feeder to the pyrolysis reactor, had to be installed in order to reach the feeding system which is positioned on the floor above the main reactor unit. This extension was built from 1-1/2-inch nominal, schedule-10S, stainless-steel pipe having a length of about 4 feet. A 2-inch gate valve was installed in the feed column just below the flange, which connects the column to the feeder, to provide the capacity to separate the feeding system from the rest of the reactor during a run for access to the feed hopper. A three-way valve is located above the gate valve so that the nitrogen purge can be directed below the gate valve when it is closed.

Data Acquisition System

A data acquisition system, originally installed by Bezama, (1983) allows on-line data from key points of the process to be automatically displayed, filtered, and stored for later off-line processing. The three main sections are illustrated in Fig. 15.

The first section consists of hardware installed on the tar-sand unit for the purpose of measuring temperatures, pressures, electrical power input, and fluid flow rates. Presently, these measurements can all be displayed on the process control panel; in addition, temperature and pressure signals can also be sent to the interface section where these analog signals are digitized and temporarily stored for further processing. The temperature and pressure measuring equipment will be discussed further in the section on temperature and bed-level control.

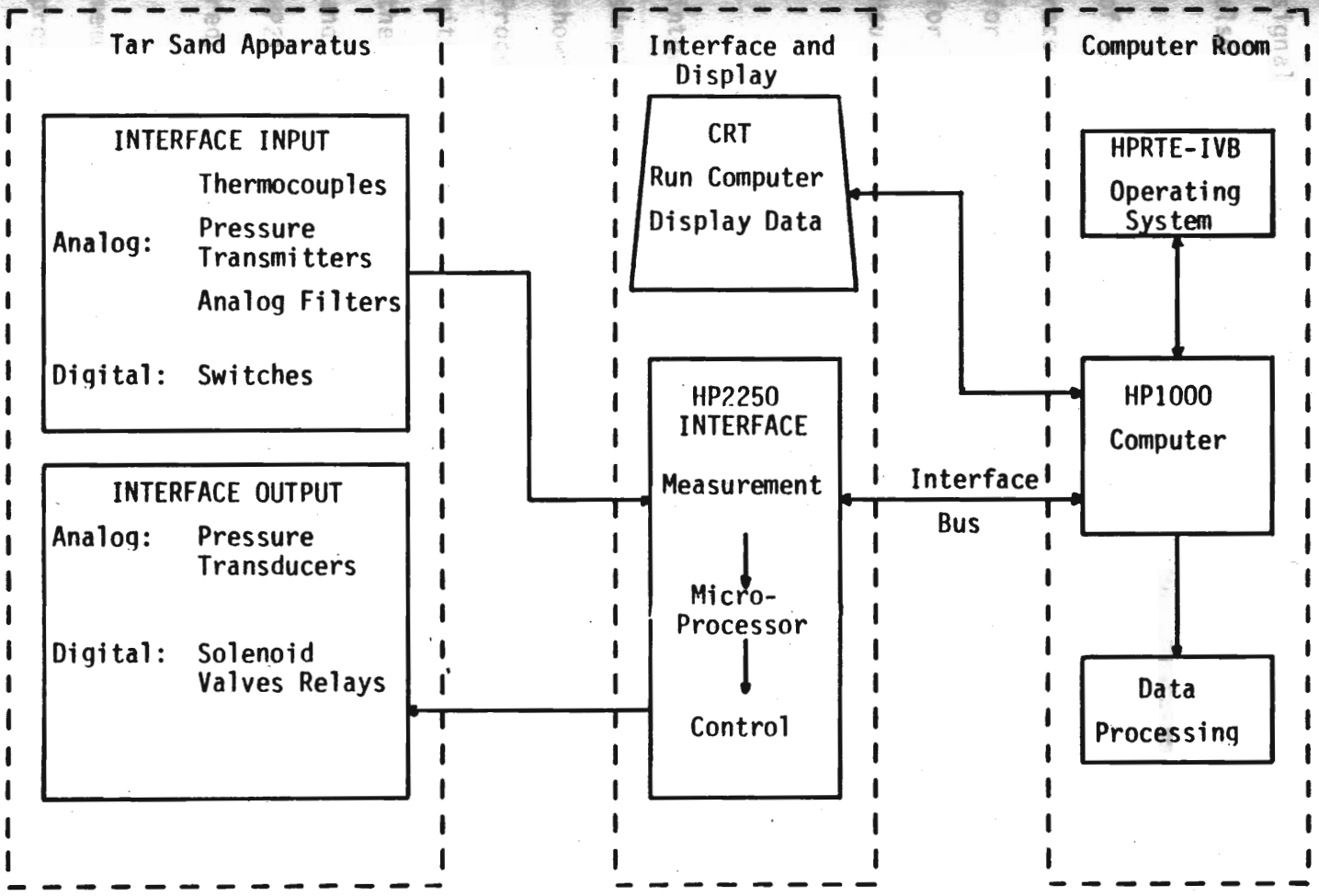


Figure 15. The data acquisition system

The second section features an HP2250-intelligent-interface unit where the analog signals from the control panel are converted to digital signals to be used for data logging and process control. This section also includes a CRT terminal used for data display and system operation.

The third section is an HP1000 computer system with various types of storage and display devices used for final processing of the data received from the interface. A brief description of the hardware system for the HP2250 interface will be followed by a discussion of the method for on-line data acquisition which was developed for the most part by Hsu (1983) and Bezama (1983).

Measurement and Control Processor

The HP2250 measurement and control unit is a real time programmable interface between the controlling and the processing system. A simplified diagram of the measurement and control processing system is shown in Fig. 16. The HP2250 system, which consists of a HP2104 Processor Unit and a HP2251 Measurement and Control Unit, communicates with the controlling computer through an interface bus (HP-IB) which is the standard method of communication between bus-compatible instruments and HP computers. The measurement and control signals are sent to the HP2250 through function cards that are chosen to meet specific requirements.

The HP2104 processor unit consists of a card frame to hold the memory card, the measurement and control interface card, and the processor card. The processor unit executes instructions programmed in MCL/50 language to provide control over the I/O function cards of the HP2251 measurement and control unit (MCI) and communicates with the host

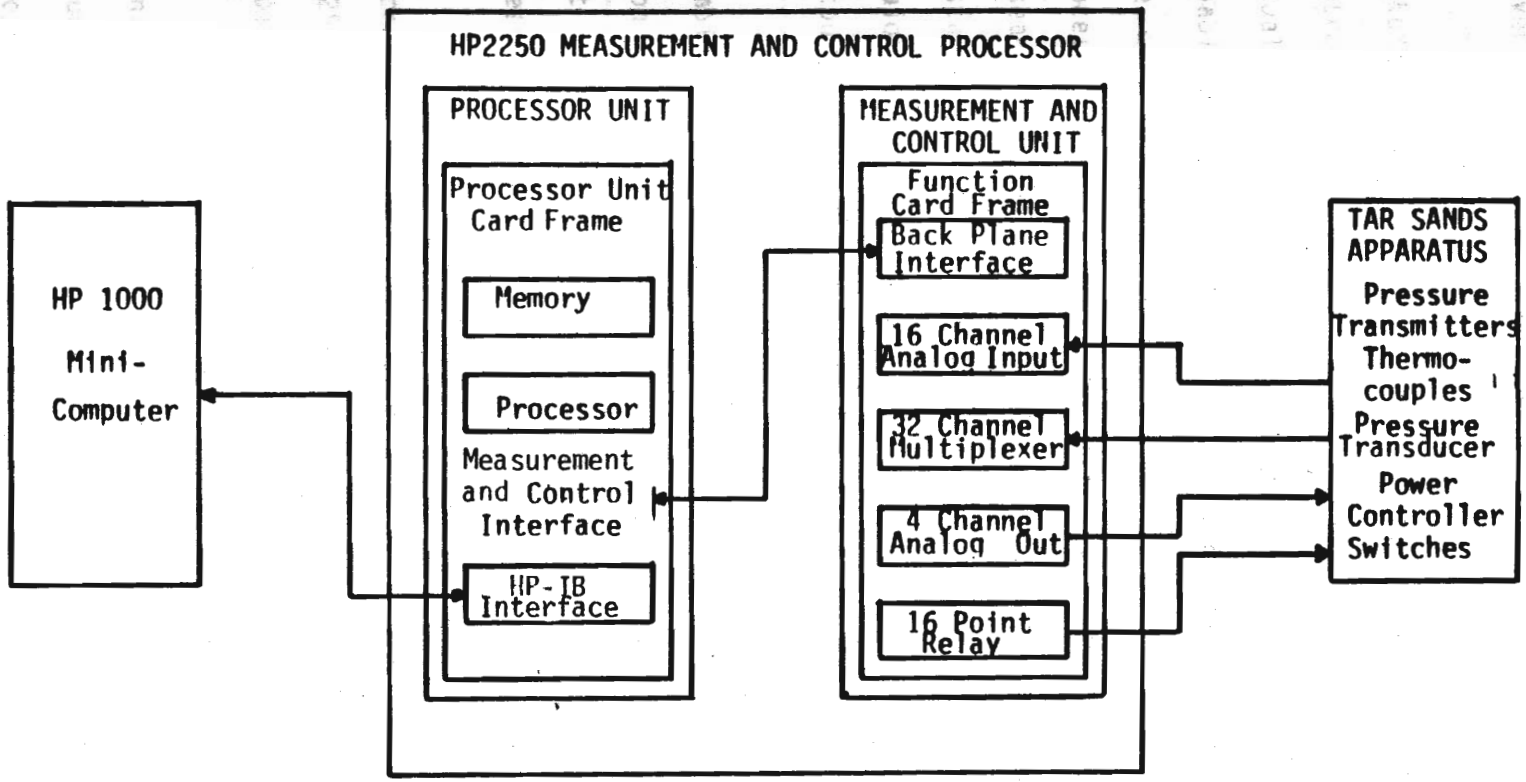


Figure 16. The measurement and control processing system

computer via the HP-IB interface to execute requests, and reduces and converts measurement and control data.

The measurement and control unit consists of a card frame and a backplane interface card which interfaces the other measurement and control function cards of the MCU to the HP2104 processor unit. Actually, a total of eight different measurement and control units, each capable of holding up to eight measurement and control function cards, are capable of being interfaced and controlled by a single HP2104 processor unit. Only four function cards are being used in this research at the present time. These cards, which were originally installed and tested by Bezama (1983), include a 16-channel high-speed analog input, a 32-channel low-level multiplexer, a 4-channel analog output and a 16-point relay output.

The HP25501A provides the basic analog input for the interface by combining high speed analog-to-digital conversion and gain amplification. This card, which features 16 addressable input channels with 14-bit resolution, is used to digitize signals coming from the four pressure transducers installed on the reactor.

The HP25503A 32-channel low-level multiplexer offers interfacing to microvolt and millivolt inputs. When used in combination with two HP25594A thermocouple reference connectors, this card provides reading capability for 30 type-k thermocouples.

The HP25510A 4-channel output card provides the digital-to-analog conversion necessary for independent analog output. The card features four programmable digital-to-analog converters and is used for proportional control of a single-phase, phase-angle-fired power

controller, which regulates the pyrolysis bed temperature, and two pressure transducers, which control the fluidized-bed levels.

Finally, the HP25514A 16-point relay output provides the switching capability of 16 open circuit voltages. The 16 channels can be programmed as independent points and are used, with installed fuses for card-overload protection, and for on/off switching of various system devices for run operation. A more detailed description of the function card capabilities is found in the appropriate Hewlett Packard hardware reference manuals and in the works of Bezama (1983) and Hsu (1983).

Data Acquisition and Processing

During a run, the HP2250 interface is programmed to sample the on-line data from 16 thermocouples, 4 pressure transmitters, 2 real-time digital inputs, an analog input, and a real-time clock. These data are stored in a buffer for up to three minutes to provide a suitable data history for the host computer. The time not used for channel sampling is used for operation of a control task that controls the upper-bed level temperature and the fluidized-bed levels, to be discussed later.

The signals for the temperature measurement are passed through signal conditioning modules installed on the low-level multiplexer card for low-pass RC filtering. Pressure measurement noise is filtered by using an analog filter, consisting of a small tank with a 0.1 second retention time, installed in the pneumatic-signal line coming from the differential pressure cells. A digital filtering routine, involving 2 different algorithms, is used to filter the digitized data for the temperature and pressure measurements. The first algorithm uses the MCL/50 command in the interface to compute the arithmetic average;

whereas, the second algorithm uses the numerical equivalent of first order lag. Data filtering for this research is discussed in greater detail by Hsu (1983).

After the data are filtered, they are stored on buffers to be transferred later to the host HP1000 computer. The data are then compressed and stored in the HP7906 MR hard-disk system. Management of these stored data is done by using several routines developed by Bezama (1983) and discussed by Hsu (1983).

Bed Level Control

Research by Wang (1983) concluded that the C_5 hydrocarbon liquid yield from the thermal processing of Utah tar sand in a fluidized bed was inversely proportional to the solids retention time. In order to control solids retention time for this research, the fluidized-bed height and the solids flow rate into the reactor would have to be regulated. Two different methods for bed level control were considered when designing the unit. The first possiblity was to install a simple overflow pipe to control the height of the bed. Other than simplicity, this choice offered no real advantages because using an overflow pipe would allow leakage of gases from one bed to the other and would not easily permit variation of bed heights for different runs. For this reason, a solids control valve, consisting of a 1/2-inch stainless-steel slide rod controlled by a pneumatic actuator, was used to control the bed level.

The instrumentation diagram showing the method for bed-level control is shown in Fig. 17. Currently, the system is capable of being controlled either by analog Foxboro Stabilog Model 40 pneumatic

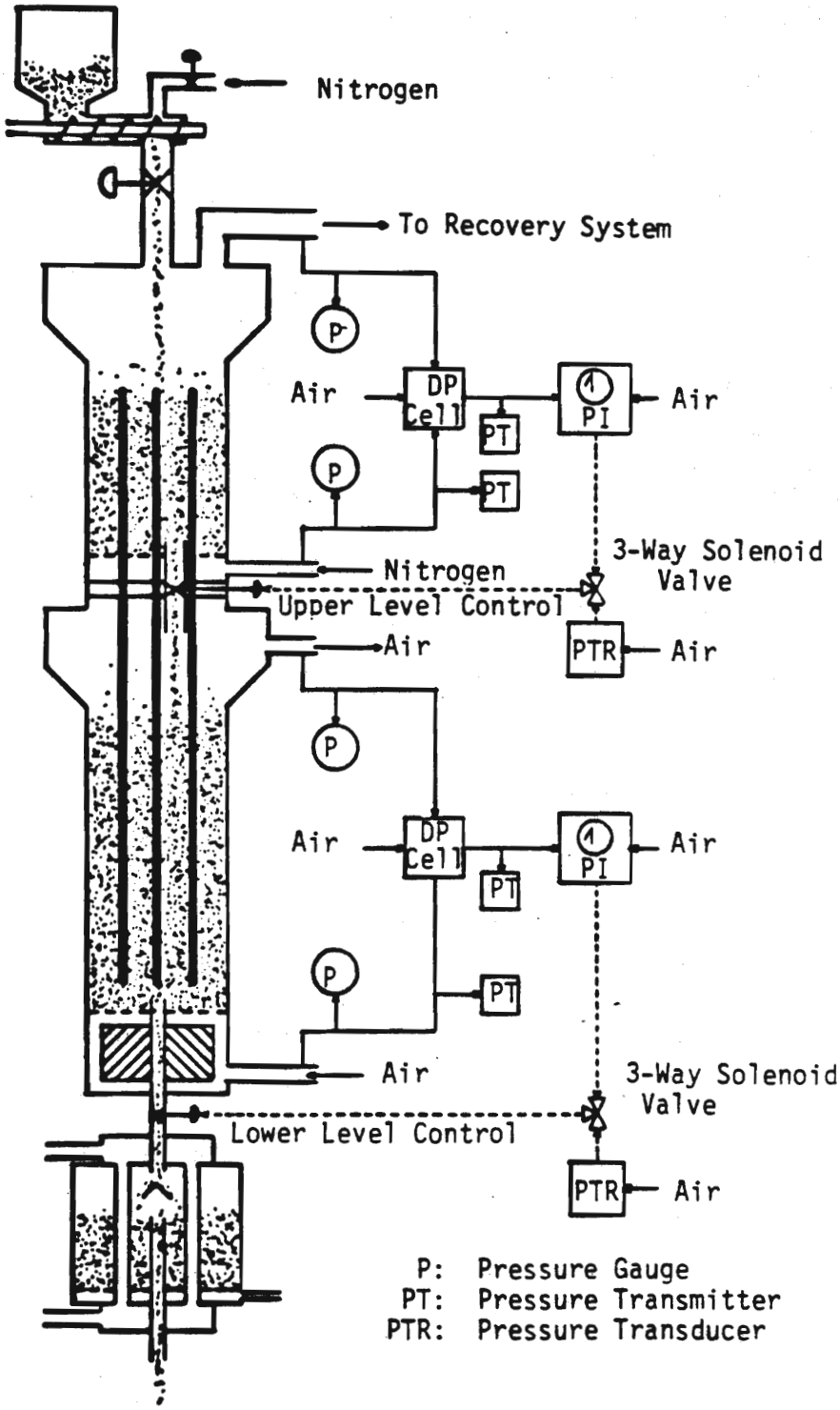


Figure 17. The instrumentation diagram for upper and lower bed-level control

controllers or by digital control using the HP2250 interface. Analog signals, corresponding to measured pressure drops across the fluidized beds, are sent to a controller, which compares the signal to the previously established set point; consequently, control signals are generated and sent to operate the pneumatic actuator, which further opens or closes the solids control valve depending on the state of the bed level with respect to the set point. Two Foxboro differential pressure cells are used to convert the measured pressure drop across each fluidized bed into standard 3 to 15 psi pneumatic signals. These pneumatic signals can be used either by the Foxboro PI controllers for analog control or by the pressure transmitters for eventual digital control. As mentioned earlier, the pneumatic signals from the DP cells are sent through an analog filter, consisting of a small tank, to reduce the noise caused by pressure fluctuations in the bed due to bubble formation.

Besides the two pressure transmitters used to measure the pneumatic signals coming from the two DP cells, two other pressure transmitters are used to measure the maximum pressure inside each reactor. All four of the Viatran Model 501 pressure transmitters were calibrated by the manufacturer; in addition, the two transmitters used to measure the pressures from the DP cells, which correspond to the pressure drops across the fluidized beds, were calibrated by Hsu (1983). The voltage signals from the pressure transmitters are then sent to four of the channels in the HP25501A function card of the interface.

Two Conoflow model T-28 pressure transducers, calibrated by the manufacturer, are used to convert the control signals, coming from the

analog-output function card of the interface, into standard 3 to 15 psi pneumatic signals for operation of the valve actuator. Two 3-way solenoid valves are used to allow for conventional proportional integral (PI) control when the valve is de-energizied.

A detailed discussion of the control strategy used to develop the Proportional Integral Derivative (PID) routine for bed level control, the gap method for solids-valve linearization, and the procedure used to tune the control algorithm is given by Hsu (1983) and is not discussed here. Performance of the bed-level control will be shown in the results section of this thesis.

Pyrolysis Temperature Control

There exists an optimum pyrolysis temperature at which the yield of C_5+ liquid is maximized, as shown by Jayakar (1979). In order to test and verify any yield dependence on the temperature of the upper pyrolysis bed, the temperature of that bed needs to be controlled. Since the heat required to maintain the upper bed at a temperature high enough to permit pyrolysis of the bitumen is supplied entirely by the three potassium heat pipes, and since previous work with laboratory-scale heat-pipe assisted thermal processing of lean tar sands demonstrated that extra energy was needed to be added to the system to allow for heat dissipation to the surroundings, it was obvious that the amount of heat being transferred by the heat pipes had to be controlled in order to control the upper-bed temperature. However, for a system where the combustion of the coked layer provided more than enough heat to maintain the desired pyrolysis temperature, a control strategy would have to be employed that would remove the excess heat being transferred

to the pyrolysis bed. For this research, the temperature in the pyrolysis bed is controlled by the amount of heat being supplied to the lower combustion bed.

Figure 18 shows a simplified diagram of the instrumentation used for upper-bed temperature control. The thermocouple immersed in the upper fluidized bed and labelled as channel 5 in the data acquisition system, is the controlled variable which is sampled by the interface. This thermocouple is one of the 16 chromel-alumel thermocouples, with probes consisting of 1/16-inconel sheaths insulated with magnesium oxide, used in this research. Other instrumentation for temperature control includes 16-gauge nichrome heating wire used for energy input to the combustion reactor, a Powerstat variable transformer for secondary power control, and a Halmar single-phase, phase-angle-fired power controller used in conjunction with the interface for automatic temperature control.

The output of the power controller, controlled between 0% and 95% of the supply voltage, is sent through the variable transformer, used here to control the maximum output voltage of the power controller and to prevent burn-out of the nichrome heating wire, and then to the heating wire wound around the combustion reactor. A double-throw relay has been installed on the apparatus to permit bypassing of the automatic power controller if necessary.

The use of the PID control algorithm, involving the time-delay behavior of the heat pipe and the tuning of the control constants for automatic temperature control, is discussed in detail by Hsu (1983). Temperature-control data for the scaled-up reactor will be discussed later in the results section of this thesis.

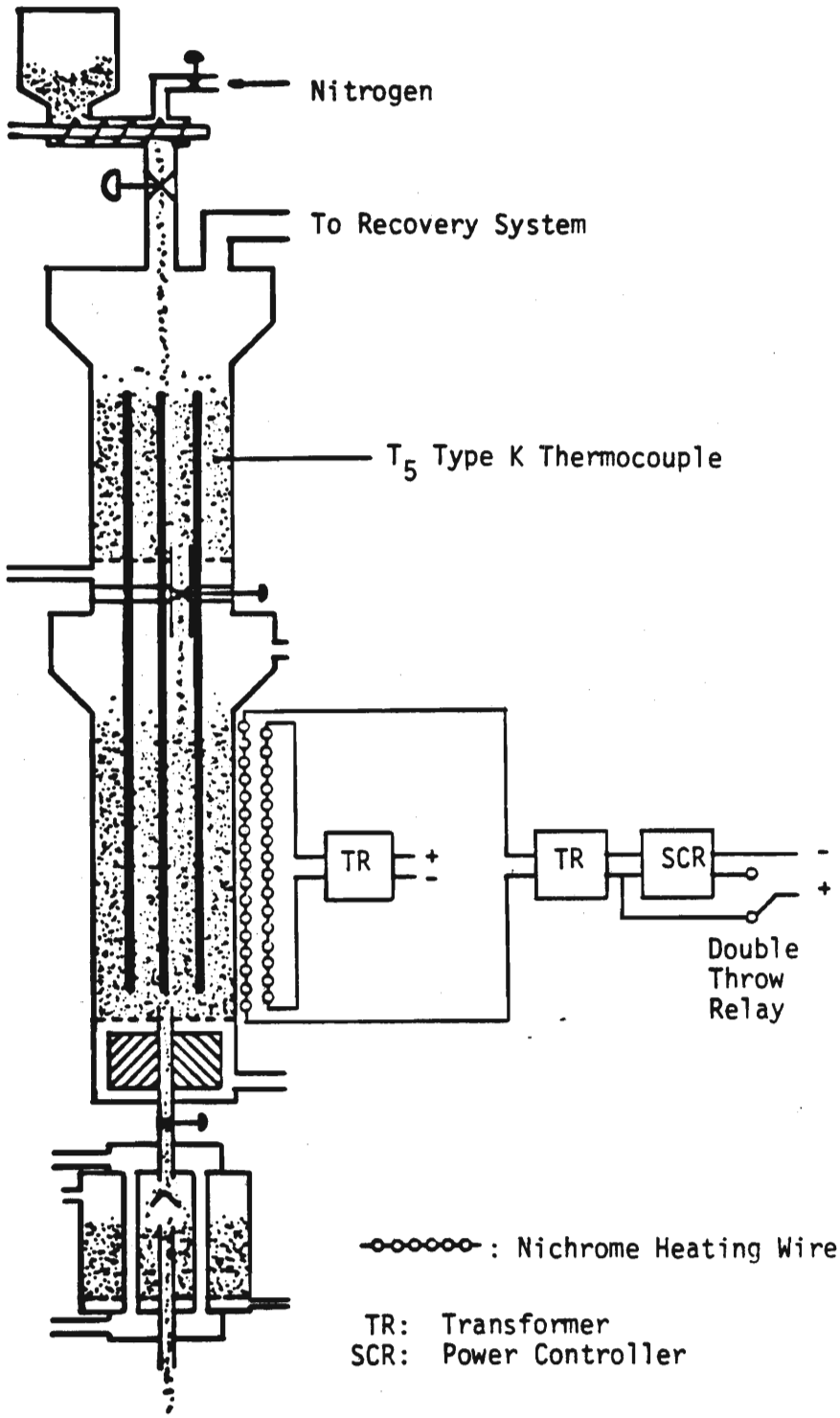


Figure 18. The instrumentation diagram for pyrolysis-bed temperature control

CHAPTER IV

EXPERIMENTAL PROCEDURE

The experimental procedure for this research can be divided into three general areas: (1) equipment calibration, (2) experimental runs, and (3) methods of analysis. The methods used in these areas will be briefly reviewed.

Equipment Calibration

The instrumentation used in connection with the digital data acquisition and control system was calibrated by Hsu (1983) and Bezama (1983); however, calibration was necessary for new flow meters, fluid-bed characteristics, and heater performance. The object of this characterization was to obtain reliable data to be used for validation of correlations and for future scale up.

Figures 19, 20, and 21 show the rotameter flow characteristics for the combustion air, fluidizing nitrogen, and purge nitrogen, respectively. The plots show actual gas flow, in cubic feet per minute, as a function of flow-meter setting. A dry volumetric meter (Singer DTM-115), calibrated at the factory, was used to acquire known measurements used for meter calibration.

To permit an experimental estimation of the average residence time of the tar sand in the pyrolysis reactor, a measurement of the mass of the fluidized bed as a function of nitrogen flow and bed pressure drop was obtained, as shown in Figure 22. This calibration was performed

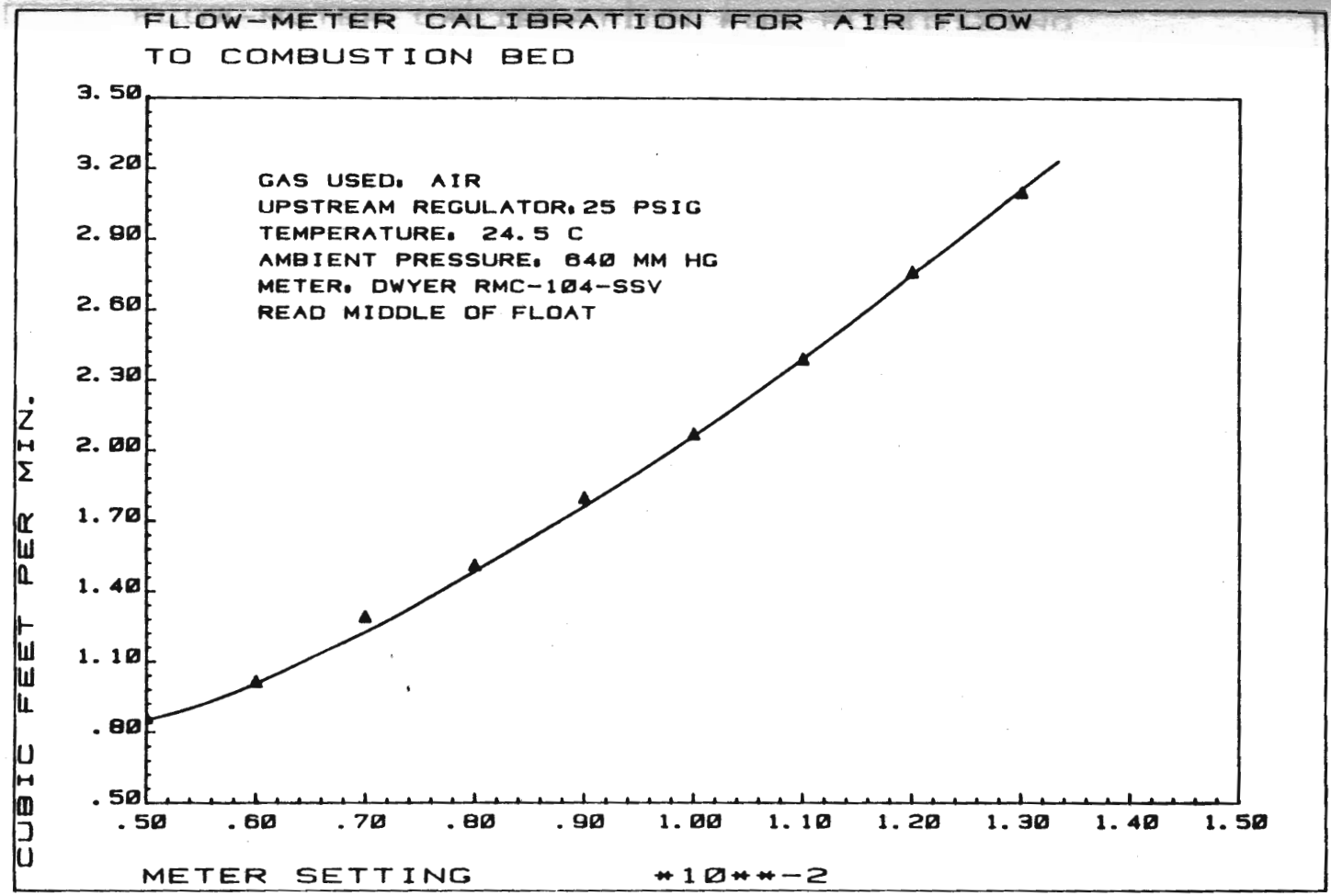


Figure 19. Flow-meter calibration for fluidizing air to the combustion bed

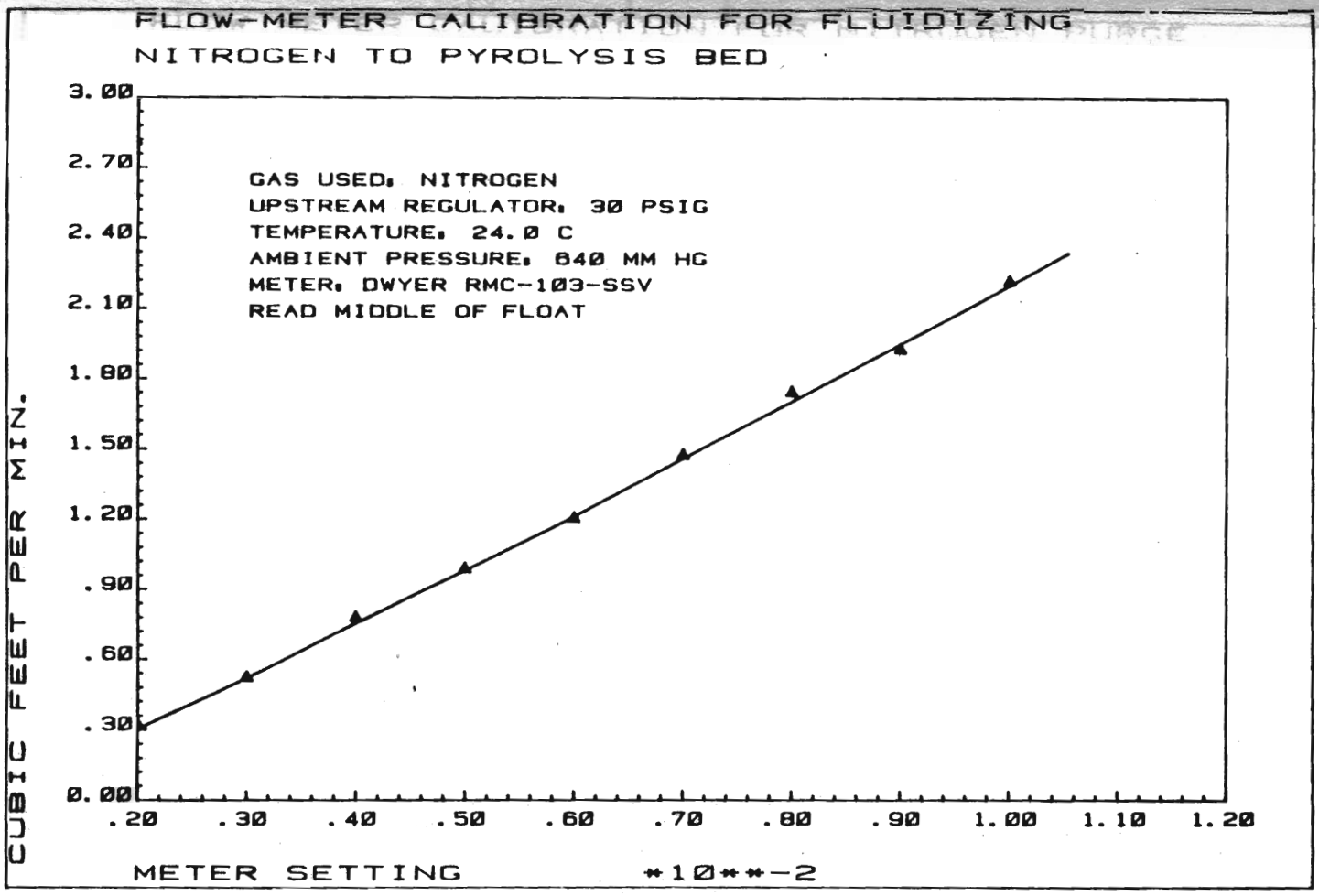


Figure 20. Flow-meter calibration for fluidizing nitrogen to the pyrolysis bed

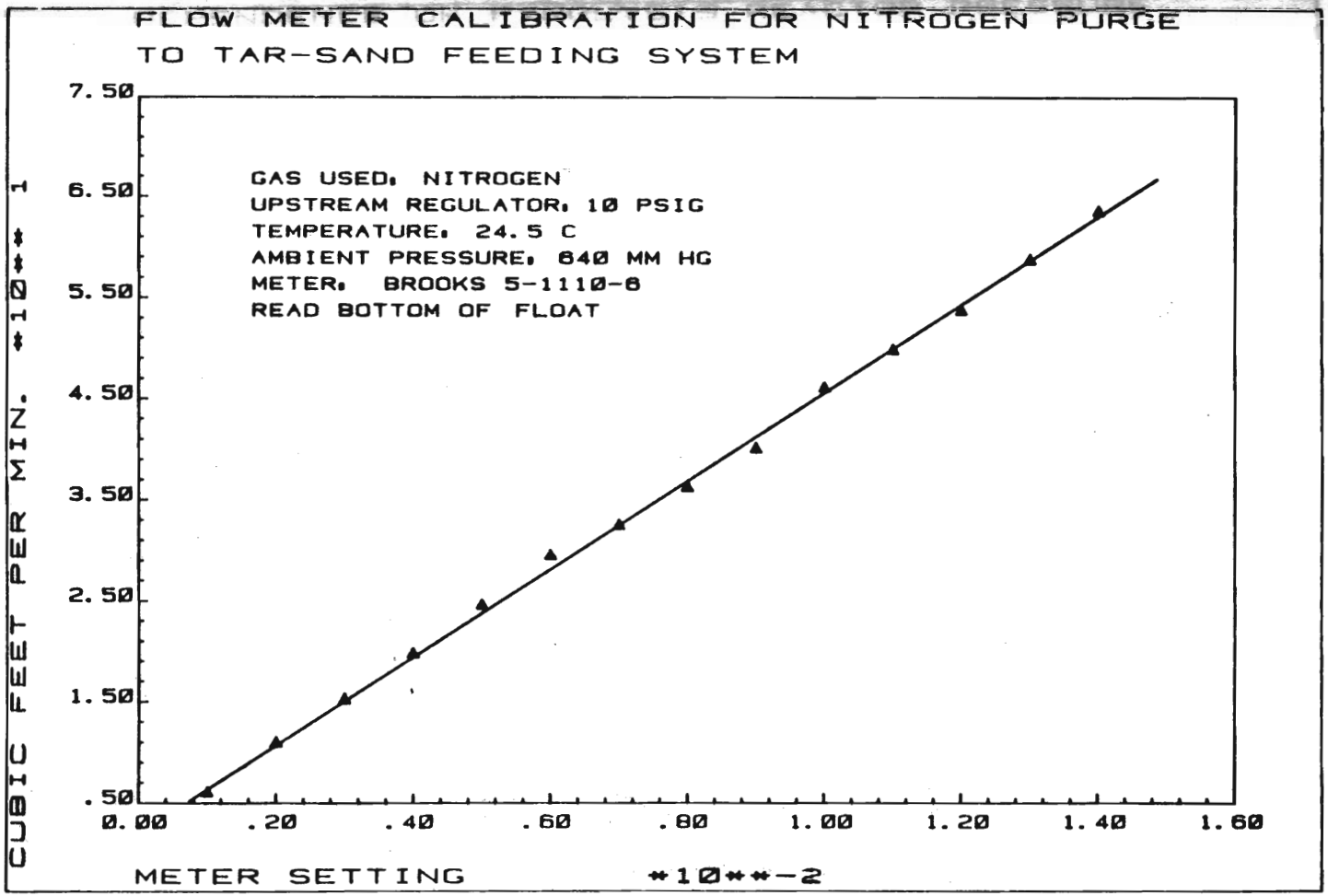


Figure 21. Flow-meter calibration for feeder-purge nitrogen

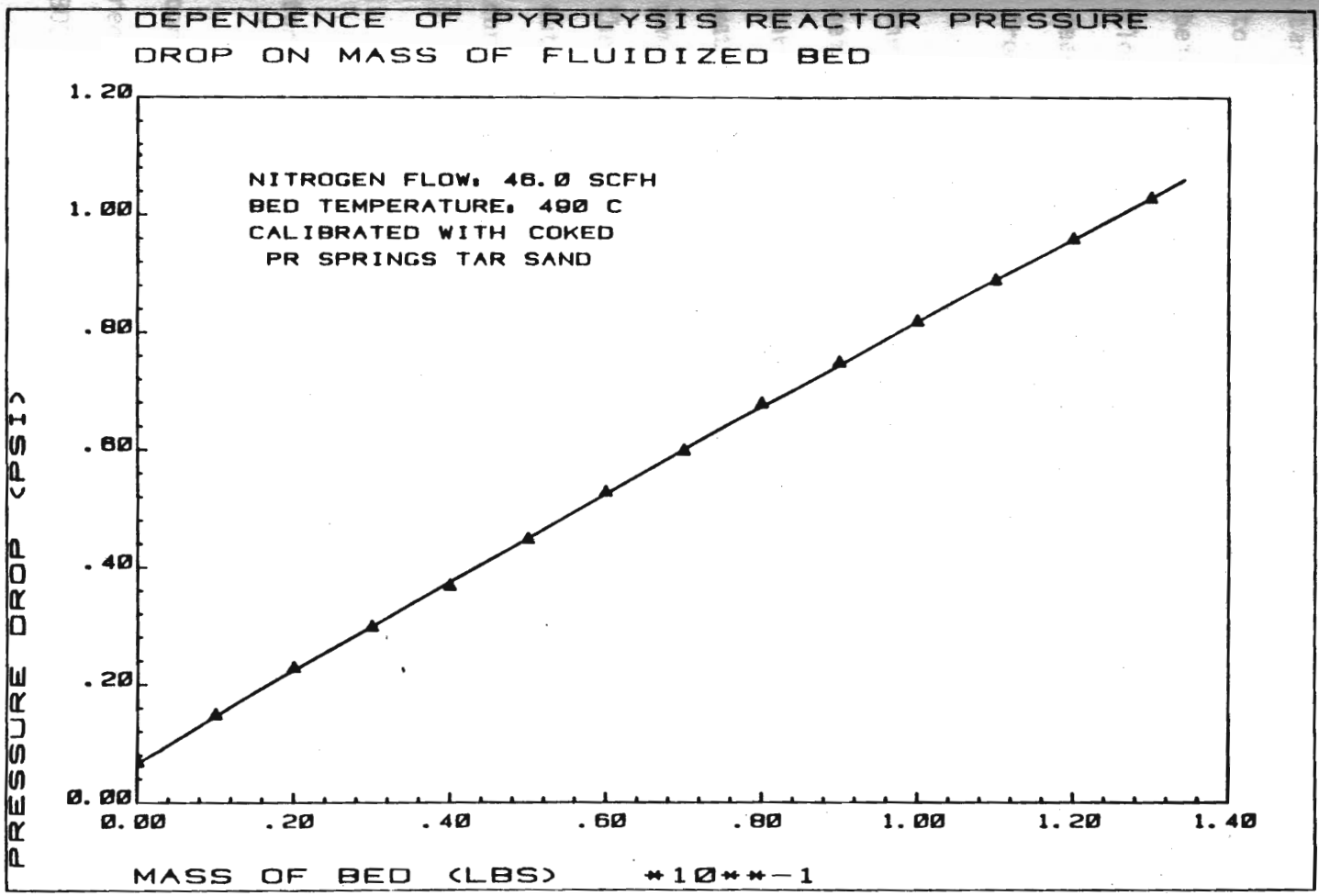


Figure 22. Pressure-drop characteristics of the fluidized coked-sand bed

under conditions similar to those of the experimental runs, using coked tar sand from the PR Springs deposit. The empty upper bed was allowed to come to a temperature of 490°C using digital PID control, the heat being solely supplied by the three heat pipes. With a constant nitrogen flow of 46.0 SCFH, one pound amounts of coked sand were added to the reactor. After the fluidized bed had come to temperature, the reading from the pressure transmitters, calibrated by Hsu (1983), was recorded.

A reliable value of the energy usage is essential for determining the applicability of a thermal process. Measurement of the energy, being supplied to the internal heating wires of the combustion bed through the SCR power controller, was accomplished with a single-phase watthour meter (General Electric Type I-50-A), which is ensured by the manufacturer to give a measurement, to within 0.5 percent accuracy, of the kilowatt hours consumed. The amount of heat being supplied by the outer combustion heating wires was determined from their constant current and potential.

Operating Procedure

This section presents the operating procedure used for making the tar sand runs.

1. Prepare the oil-collection equipment by cleaning the two condensers and mist collector with acetone. After this equipment is dry, record their weights from the triple-beam balance (Henry Tromner 20-803).

2. Prepare the solids separation system by removing the fines collected from the previous run. This is accomplished by using a

reverse flow of compressed air through the sintered-steel element. The filtering section is then reassembled with a new gasket.

3. Check all electrical connections and insulation to see that they are in proper position.

4. With clean sand in the hopper, and fluidizing air flowing in both beds at the appropriate rotameter settings, allow both beds to fill to the desired levels, determined by the set points of the pneumatic controllers.

5. Turn on the heaters to the combustion bed, product exit line, and solids-filter section. Record the amperage reading for each heater.

6. After vertical expansion of the unit, due to heating, has stopped, assemble the two condensers and the mist-collection unit, and check for any leaks.

7. Turn on the HP-IB extender, HP2250 Interface, control terminal, and the pressure transducers, and pressure transmitters. Run the process-control program (TSAND) and the data acquisition subprogram (STORE), recording the appropriate input used.

8. Begin digital control of the temperature and fluidized-bed levels. Perform any tuning necessary to improve the bed-level control, and allow the upper bed to reach the set temperature with a constant feed of clean sand flowing through the system.

9. After the system is at the desired control temperature, switch the upper fluidizing flow and the upper-valve purge to nitrogen. Start the nitrogen feed purge and cooling-water flow.

10. Switch the control system to the data storage mode and load the hopper with the sized and weighed tar-sand feed.

11. Record the time when tar sand feeding is initiated. Record the kilowatt hour reading for the inner-combustion heater and the amperage for the outer-combustion heater.

12. Monitor the run for any significant event dealing with bed-level and temperature control, nitrogen supply, solids flow, or any other possible problems.

13. After the desired amount of tar sand has been processed, record the time and energy reading for the inner-combustion heater. Shut off all heaters and allow the beds to cool below 300°C. Collect the oil that has drained into the collection vessels below the condensers.

14. Switch the control program to the end-run mode and input all necessary comments to be stored with the data from that run. Shut down the data acquisition and control equipment.

15. Collect solid samples from the pyrolysis and combustion beds. Dissemble the condensers and mist collector, and record their weights on the triple-beam balance.

16. Shut off all power and flows.

The major difficulties that were observed during a run included a gradual build up of pressure in the upper reactor due to cake formation on the sintered-steel filter, mechanical problems associated with the upper-solids-control valve, and occasional bridging of the tar sand in the feed hopper.

As long as the upper valve was free of any sand, which would cause binding between the valve rod and housing, it would function quite adequately. Occasionally, when the upper valve would partially open,

sand-flow would momentarily plug, causing the valve to open further. The valve would then continue to open until the slight plug was displayed, allowing unimpeded flow through the, now, fully opened valve. For this particular valve, the sand flow is then nearly 1/2 lb/sec and by the time the valve is closed again, significant bed loss can occur. However, no other serious plugging problems for the two solids-control valves were ever observed.

Regulator pressures on the upstream side of all gas flows were kept at the pressures used for meter calibration to insure a fairly accurate measurement of the flows.

Processing of Tar Sands

Experimental runs for this research were made with four different types of Utah tar sands. Two different samples from the PR Spring deposit were tested; furthermore, runs were made using tar sands from the Sunnyside and Whiterock deposit. The samples from the PR Spring and Whiterock deposits were obtained, with permission, from the site of the Enercor pilot-plant in North Salt Lake. The Sunnyside tar sand was obtained from the Department of Fuels Engineering at the University of Utah. All tar sand samples were received already sized to chunks less than 3/4 inch. The tar sand was then ground further in a jaw crusher (Sturtevant Type 2x6). The crushed sand was sized in an 18-inch vibrating separator (Sweco Model S18S333) to obtain a feed material of -14 + 24 Tyler Mesh. Preparation of the tar sand feed was tedious and time-consuming with Sunnyside tar sand being the most difficult to work with, because of its sticky nature.

The data from experimental runs were automatically logged by the HP2250 interface system. Samples of on-line data were recorded on a real-time basis to monitor 4 pressure transmitters and 16 thermocouples. The temperature and pressure data are presented in Appendix C.

Runs With Sunnyside Tar Sands

The initial runs in this research used Sunnyside tar sand. Approximately 15 runs were attempted; unfortunately, some of the earlier runs were aborted because of difficulties encountered with the newly assembled equipment. The major problems encountered with the Sunnyside runs were bridging problems in the feeding system, plugging of the cyclone and the original exit line, mechanical failure of the screw-feed system, and inadequate performance of the upper solids-control valve.

The bridging problem of the feed was eliminated by changing the angle of repose in the hopper and installing a gate valve, which allowed run-time access to the hopper without any loss of oil vapors. A view port and a manual stirring device were installed in the lid of the hopper to further minimize any disruptions of the feed rate. Since the runs with the scaled-up unit would process up to 12 times more feed than the original reactor, the excess loading of fines in the exit line and the cyclone, which was located directly below the filter, caused serious plugging problems, This problem was solved by replacing the original 1/2-inch exit line with a line fabricated from 2-inch-nominal pipe sections. The cyclone, which was located directly under the solids-filter section, was also eliminated because of continual plugging problems. Once these modifications were made, no further plugging occurred in the product recovery section, although significant pressure

build-up would occasionally occur because of cake growth on the sintered-steel filter element.

With the above changes made, 6 runs were made with Sunnyside feed in order to study liquid-crude yields as a function of the pyrolysis-bed temperature and the average solids-retention time. For the Sunnyside runs only, some spent sand was mixed with the tar sand to enhance feeder performance. Runs were made at three different pyrolysis temperatures and at two different bed hold-ups in the pyrolysis reactor. Some possible problems still occurring with the Sunnyside runs were inadequate collection of oil mist by the mist collector, and possible loss of oil vapors through the upper-solids control valve, which was modified after these runs were made.

Runs With Whiterock Tar Sands

Only two runs were made to process tar sand from the Whiterock deposit. This sample seemed to be significantly less consolidated than the Sunnyside tar sand, and resulted in a much increased rate of fines accumulation on the solids filter in the product exit line. A more effective dust-elimination system will have to be installed in order to study this tar sand further. A possible improvement would be to add an internal cyclone to the pyrolysis reactor, with a dip leg extending down into the fluid bed. This modification would greatly reduce the amount of fines accumulation on the filter, and allow longer, uninterrupted runs.

Runs With PR Springs Tar Sands

Two different tar sands from the PR Springs deposit were studied, most of the runs being made on the richest of the two samples. Ten

successful runs were made to again study the effect of pyrolysis temperature and average solids residence time. For these experiments, the upper solids control valve had been modified by installing a high-temperature packing unit around the valve rod to eliminate any leakage of oil vapors at that point. The mechanical performance of the valve had also been improved because of a gas purge that had been installed to provide a gas bearing between the valve rod and the housing. The performance of the mist collector had also been enhanced by replacing the pyrex-wool reverse-flow depth-type filter element with alternate layers of spun-bonded polyester air-filter pads and cellulose fiber. Gas-chromatographic analyses were performed on the gases leaving the mist collector to verify adequate operation.

CHAPTER V

RESULTS AND DISCUSSION

The experiments performed in this research yielded information regarding important aspects of the University of Utah Thermal Process. The main areas of the process to be discussed are controllability, energy usage and thermal performance, and physical properties and yield of the liquid product. Data from 16 successful tar-sand runs are used to study any trends of oil yield or product quality with respect to the operating parameters such as pyrolysis bed temperature and average solids-residence time in the pyrolysis reactor.

Process Control

All tar-sand runs experienced adequate control to allow process variable study. Digital PID control was used for the tar-sand runs with proportional gains of 50 and 80, reset times of 40 and 40 seconds, and derivative times of 0 and 0 seconds, for the combustion bed and pyrolysis bed, respectively. Temperature control was also accomplished with digital PID control using a proportional gain of 250, a reset time of 7 seconds, and a derivative time of 1 second. Further discussion of this control strategy, and its use in a two-staged fluidized-bed reactor is given by Hsu (1983). The most important area of control, for this process, is the bed-level control of the pyrolysis reactor. Poor upper-bed level control introduces significant disturbances to both temperature control and lower-bed level control. The temperature and pressure

histories of run number 41 (FILE41), as shown in Figs. 23 and 24, illustrate the disturbance to the process system caused by poor upper-bed control. For this run, the poor control was attributed to a bad setting of the valve, which regulates the pressure signal coming from the upper differential-pressure cell to the pressure transmitter. The limiting factor for control of this process, on a laboratory level, is the flowability of the coked sand in the upper bed. Sunnyside tar sand flows better in a coked form than does coked tar sands from PR Spring or Whiterocks, because of less fines formation during the coking process. When there is an excess of fines in the coked bed, and when the coked sand does not flow freely, hold-up occurs in the sand downcomers from the pyrolysis reactor. The upper valve will then open completely, because the upper-bed pressure drop increases above the set point, and the upper bed will continue to fill with sand until the plug in the downcomer is dislodged. The valve, now being completely open, allows sand flow at a very high rate, which causes excessive bed loss in the pyrolysis reactor, before the valve can close again. The lower bed has now received a large amount of cool coked sand, causing a large overshoot of the lower bed-level set point. This disturbance also causes a cooling of the lower bed, resulting in decreased heat transfer through the heat pipes to the upper bed. The temperature of the upper fluidized bed then drops below its set point because of incoming cool feed and decreased heat transfer from the lower bed. This type of process disturbance will probably decrease when the process is scaled up to a pilot-plant or commercial scale, where solids flow will be more consistent.

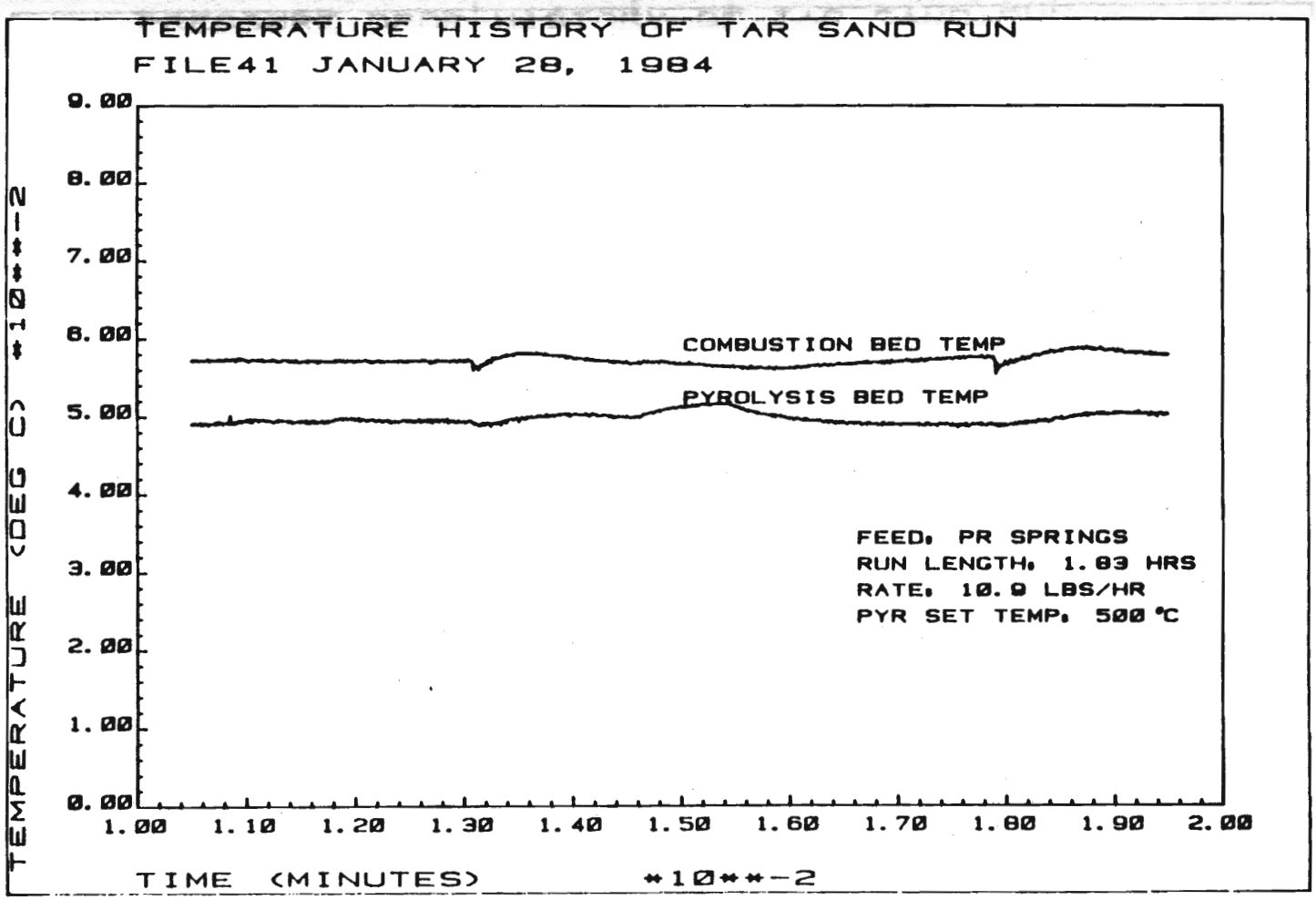


Figure 23. Temperature history of tar sand run recorded as FILE41

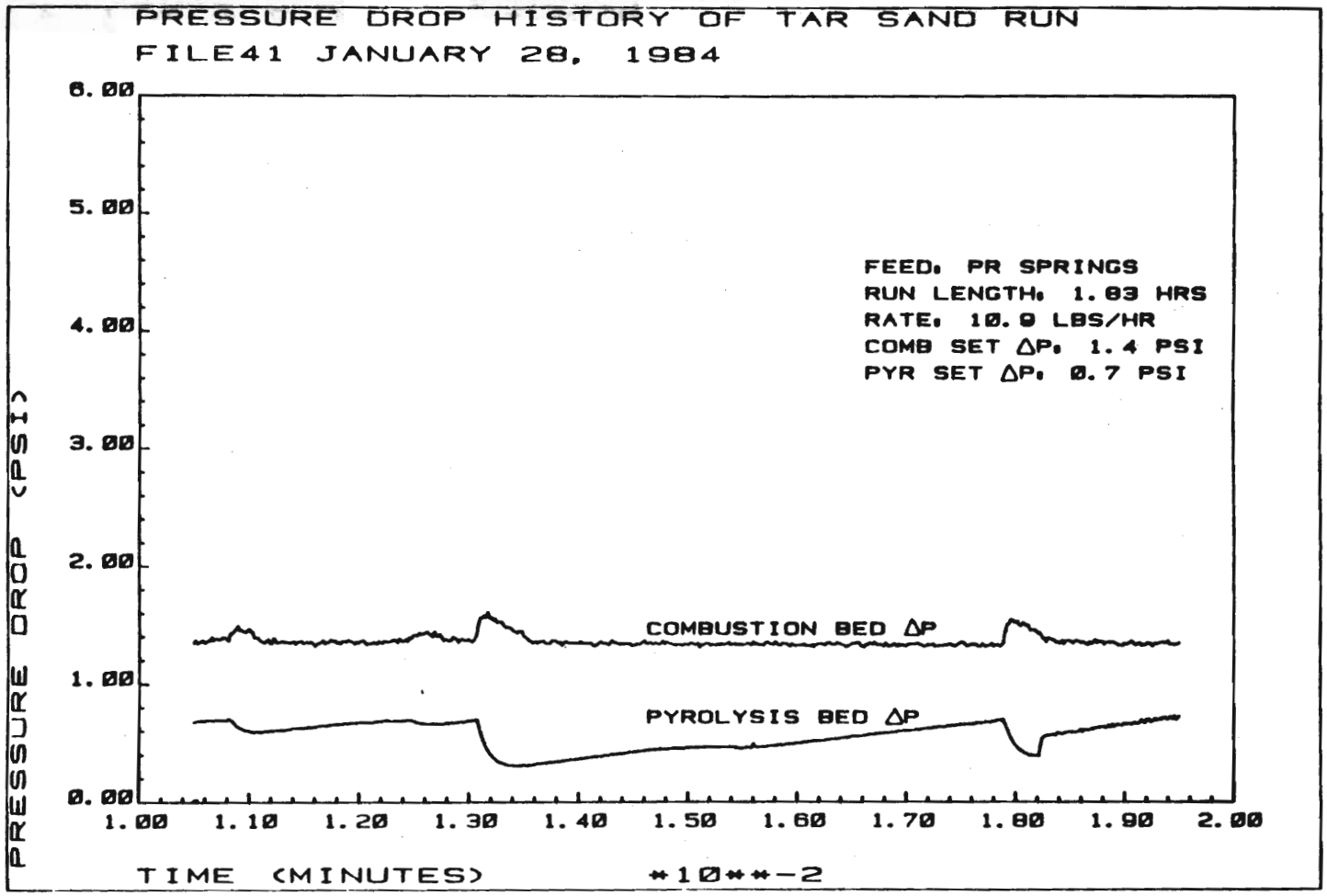


Figure 24. Pressure-drop history of tar sand run recorded as FILE41

Studies of Variables

The major process variables associated with the University of Utah Thermal Process are (1) average solids residence time in the pyrolysis reactor, (2) average gas residence time in the pyrolysis reactor, (3) fluidized-bed temperature of the pyrolysis reactor, (4) fluidizing-gas composition in the pyrolysis reactor, and (5) bitumen and feed composition. Runs, using Sunnyside feed, provided data for a study of pyrolysis-bed temperature; whereas, runs, using PR Spring tar sand, provided data for a study of pyrolysis-bed temperature and average solids residence time. Average gas residence time and fluidizing gas composition are assumed to have less of an effect on liquid yield, and the investigation of these process variables is left to future research.

Figure 25 shows the dependence of liquid-crude yield on pyrolysis-bed temperature. All of the runs shown in Fig. 25 used an average solids residence time of about 48 minutes, a flow of 70 SCFH air to fluidize the combustion bed, and a flow of 46 SCFH nitrogen to fluidize the pyrolysis bed. Other process variables, except pyrolysis-bed temperature, were kept similar and can be found in Table 8 in Appendix A. It can be seen that the liquid yield increases as the temperature of pyrolysis increases. This result agrees with Jayakar (1979) and Wang (1983). No runs were made with an upper-bed temperature above 500°C, so the optimum pyrolysis temperature cannot be reported. One of the main reasons that the liquid yields for Sunnyside and Whiterocks tar sands appear to be lower than those for PR Spring tar sands is that the mist-collection system was not removing all of the oil mist for the runs with the former two tar sands. The liquid-yield data for the PR Spring tar

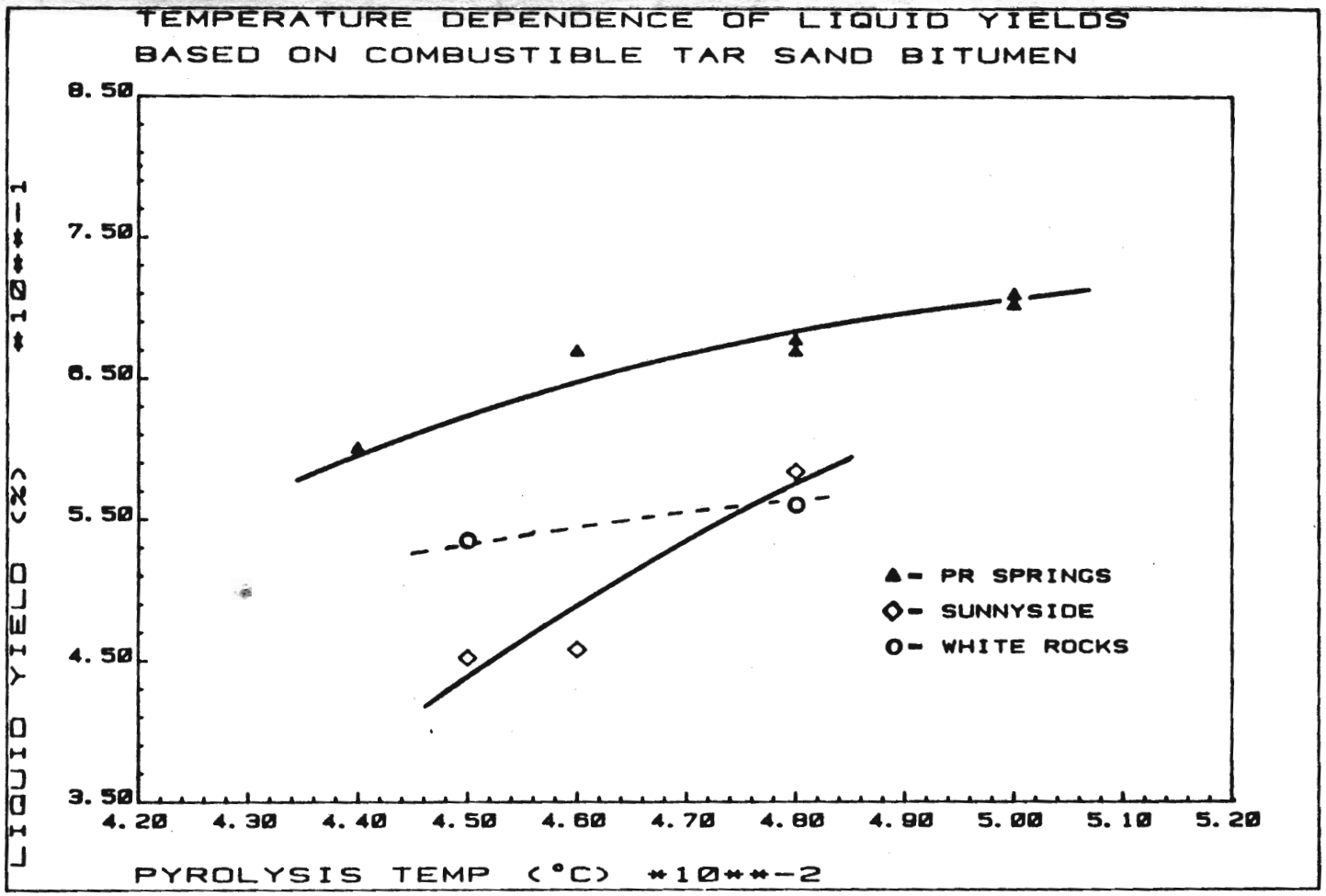


Figure 25. Dependence of liquid-crude yield on pyrolysis-bed temperature

sand is more reliable since a better mist-collection system had been installed and since the gas was sampled in an effort to achieve a complete accounting for all the material. The accounting of bitumen for the PR Spring runs ranged from 86% to 93%. Possible hydrocarbon loss for these experiments is attributed to (1) leakage of oil vapors from the pyrolysis reactor through the upper solids-control valve into the combustion reactor, (2) leakage of oil vapors from the pyrolysis reactor and product-recovery system into the laboratory, (3) analysis error in bitumen and coke content, (4) condensation of oil vapors on parts of the system which were not accounted for, and (5) other unknown reasons. However, any relative product loss that occurs on this equipment, which is capable of producing 0.14 to 0.28 barrels per day of synthetic crude form lean tar sands, would be greatly reduced for a properly designed, 1000 barrel per day, pilot plant.

Figure 26 represents coke and gas yields and their dependence on pyrolysis-bed temperature for PR Spring tar sand. Gas yields generally tend to increase with increasing pyrolysis-bed temperature; whereas, coke yield shows a general decrease. The high coke-yield values for the lower temperature runs can be attributed to incomplete bitumen pyrolysis, while at higher temperatures, a further increase in temperature results in enhanced cracking of the bitumen and less coke formation. The rise in gas yield at the higher pyrolysis temperatures is due, in part, to an increased amount of bitumen that gets cracked to a lighter hydrocarbon product.

The dependence of liquid-crude yield on average solids residence time is shown in Figure 27. The tar-sand runs were made at a pyrolysis-

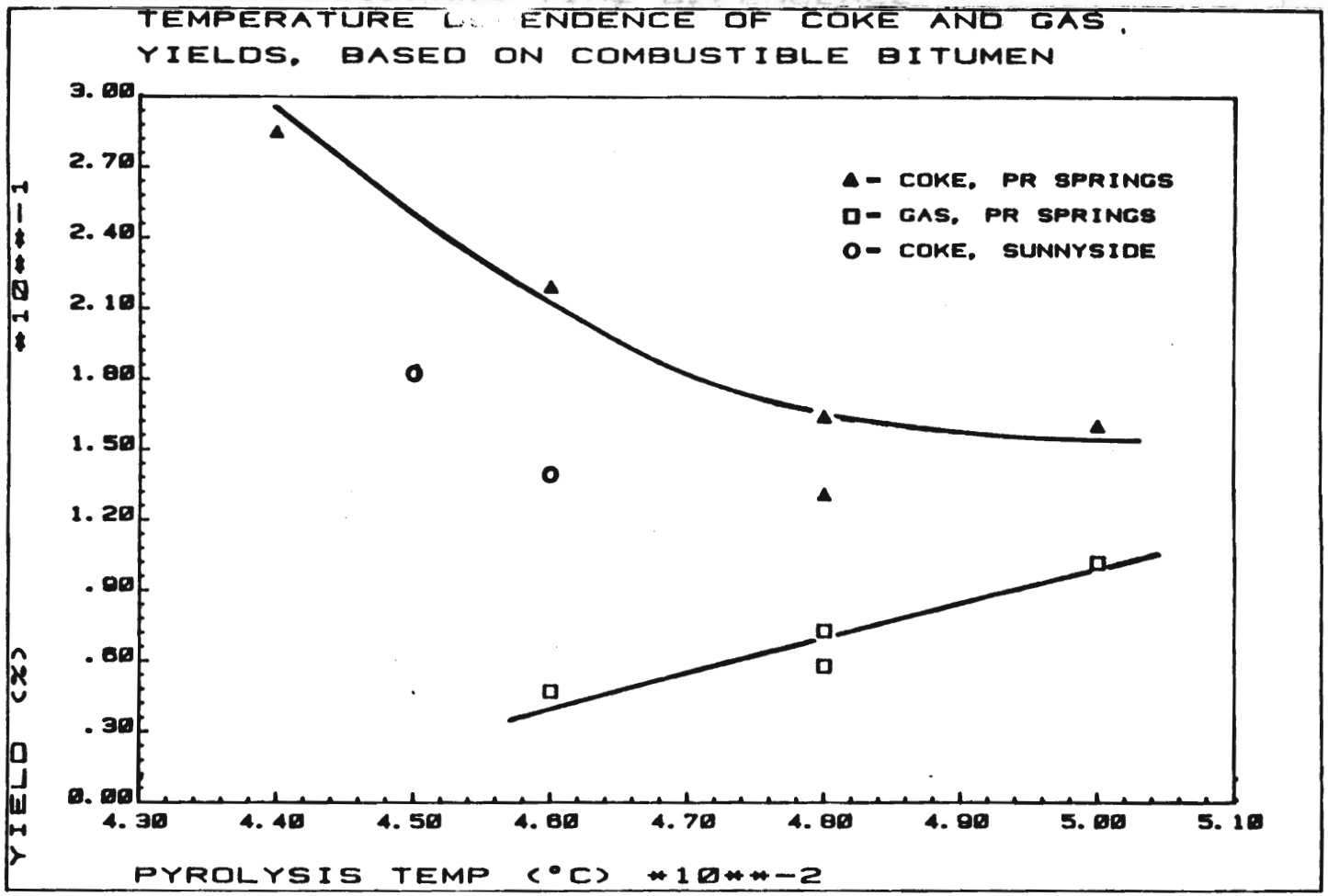


Figure 26. Dependence of coke and gas yields on pyrolysis-bed temperature

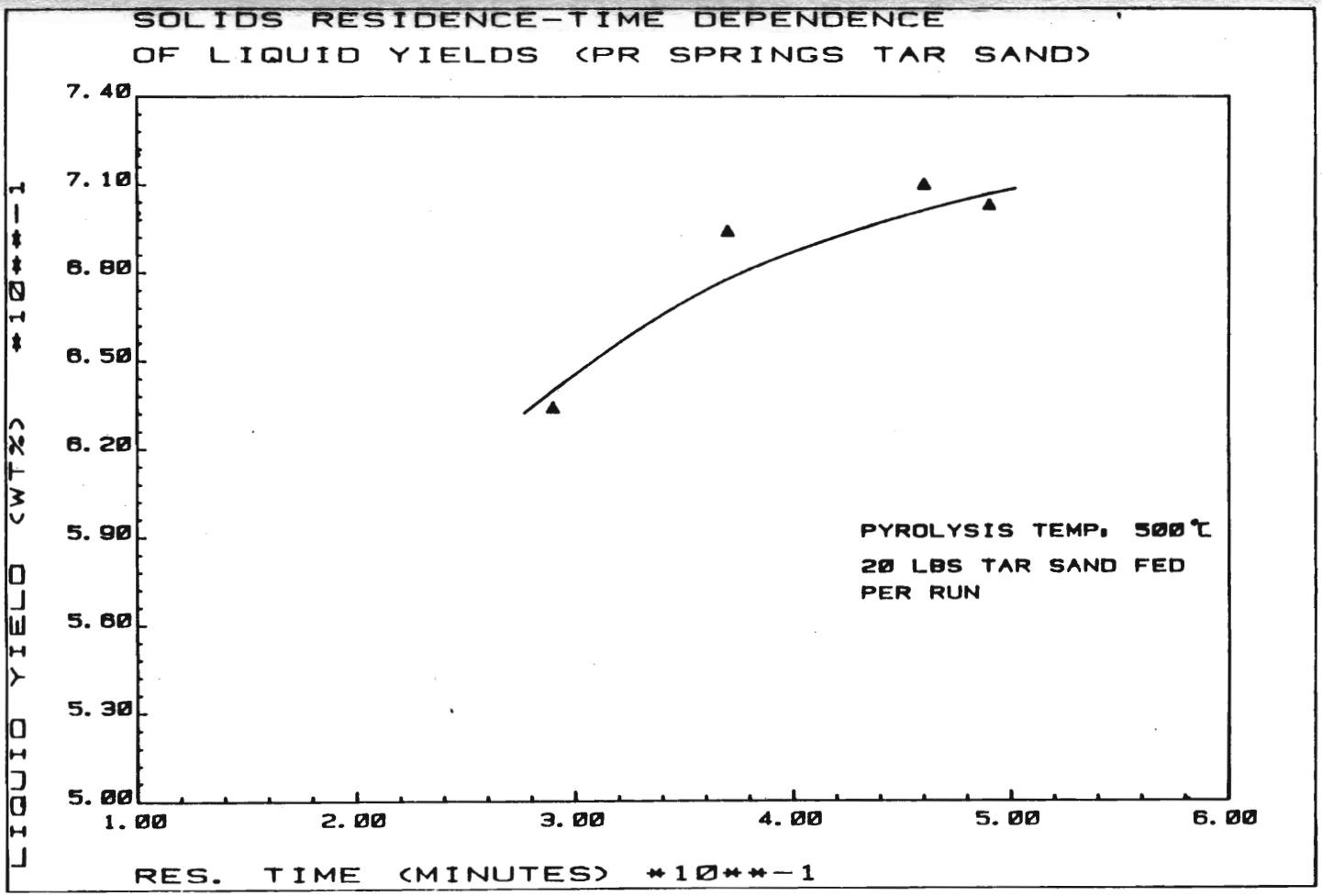


Figure 27. Dependence of liquid-crude yields on average solids-retention time

bed temperature of 500°C, with fluidizing gas and air flows comparable to the runs in Fig. 25. Other data on the run conditions are given in Table 8 of Appendix A. The results from the solids-residence study indicate that liquid yield decreases as the average solids residence time of the tar sand decreases. The high yield of 71% was obtained with an average residence time of 48 minutes for the tar sand feed, with the liquid yield decreasing to 68% for a residence time of 29 minutes. This result is in disagreement with the results of Jayakar (1979), Vankatesan (1980), and Wang (1983), wherein residence times ranged from 17 to 34 minutes. However, more runs with the scaled- up reactor would have to be made in order to formulate any concrete conclusions regarding retention time of the solids.

Product Properties

Determinations of viscosities, densities, elemental compositions, and simulated distillations were made to give an indication of the physical characteristics of the liquid products. The oil collected from the two condensers was kept separate from the oil that drained from the mist collector during a run. The reported properties of the products should be fairly reliable since none of the oils were extracted by solvents, which was the method used in previous thermal-processing studies of lean tar sands at the University of Utah. The synthetic crude was collected immediately after each run, and the product was stored in glass bottles until the analyses were made. Tables 9, 10, and 11, of Appendix A, contain all of the data from the analyses made.

Figure 28 shows that the API gravities of the oils range from 17 to 21 degrees, with no clear trend with respect to pyrolysis temperature

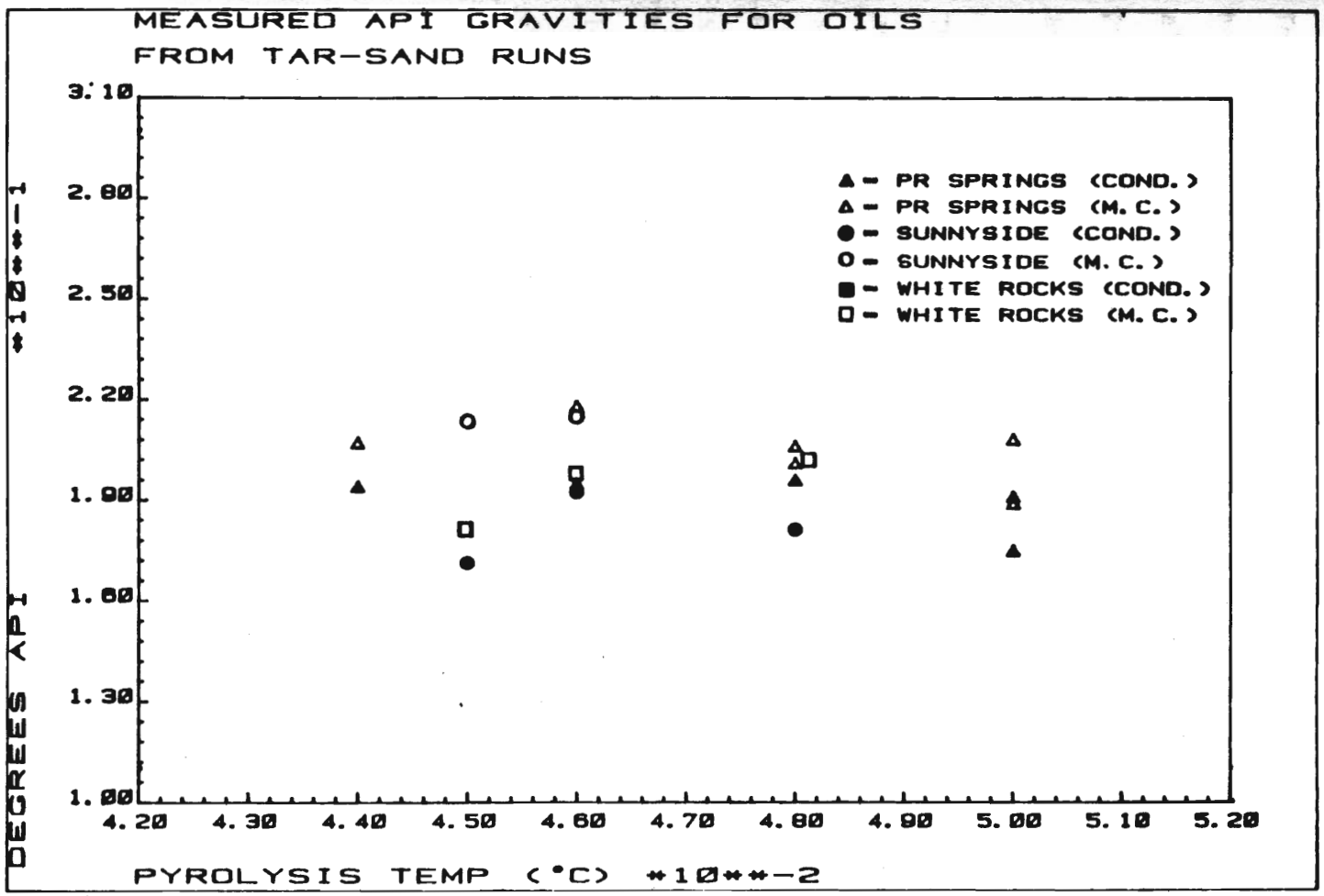


Figure 28. Measured API gravities of thermally recovered oils

apparent. However, the oil collected from the condensers is generally more dense, by 1 to 2 degrees API, than the oil from the mist collector. This range of densities places the liquid product on the border of the heavy and medium-heavy petroleum density classification (Neumann, et al., 1981). The fact that the API gravities of the products are below 20 degrees, indicates that much of the oil is of naphthene or napthene-aromatic base.

Some of the measured viscosities, at 20°C, are shown in Fig. 29. The viscosities shown range from 500 to over 2000 cp which is considerably higher than the viscosities reported from earlier work done on the smaller reactor. Again, it must be noted that solvent extraction was used in earlier work to collect the oil from the product recovery system. This could affect the resultant product viscosities. A general trend of decreasing viscosity as reaction temperature is increased can be observed, which can be explained by assuming that enhanced cracking of the more viscous components of the liquid crude occurs at the higher reacting temperatures. It can also be noted that the mist-collector oil is considerably less viscous than the condenser oil.

The C/H ratios of the product oil ranged from 0.63 to 0.77, with a higher C/H ratio for Whiterocks product and a lower C/H ratio for PR Spring products. The data for elemental analysis is found in Table 12 of Appendix A. There seemed to be no observable trends relating C/H ratio to temperature of pyrolysis. The observed values do agree with the values obtained by Jayakar (1979) and Weeks (1977). For the PR Spring experiments, the C/H ratios do not vary significantly between the mist-collector oil and the condenser oil.

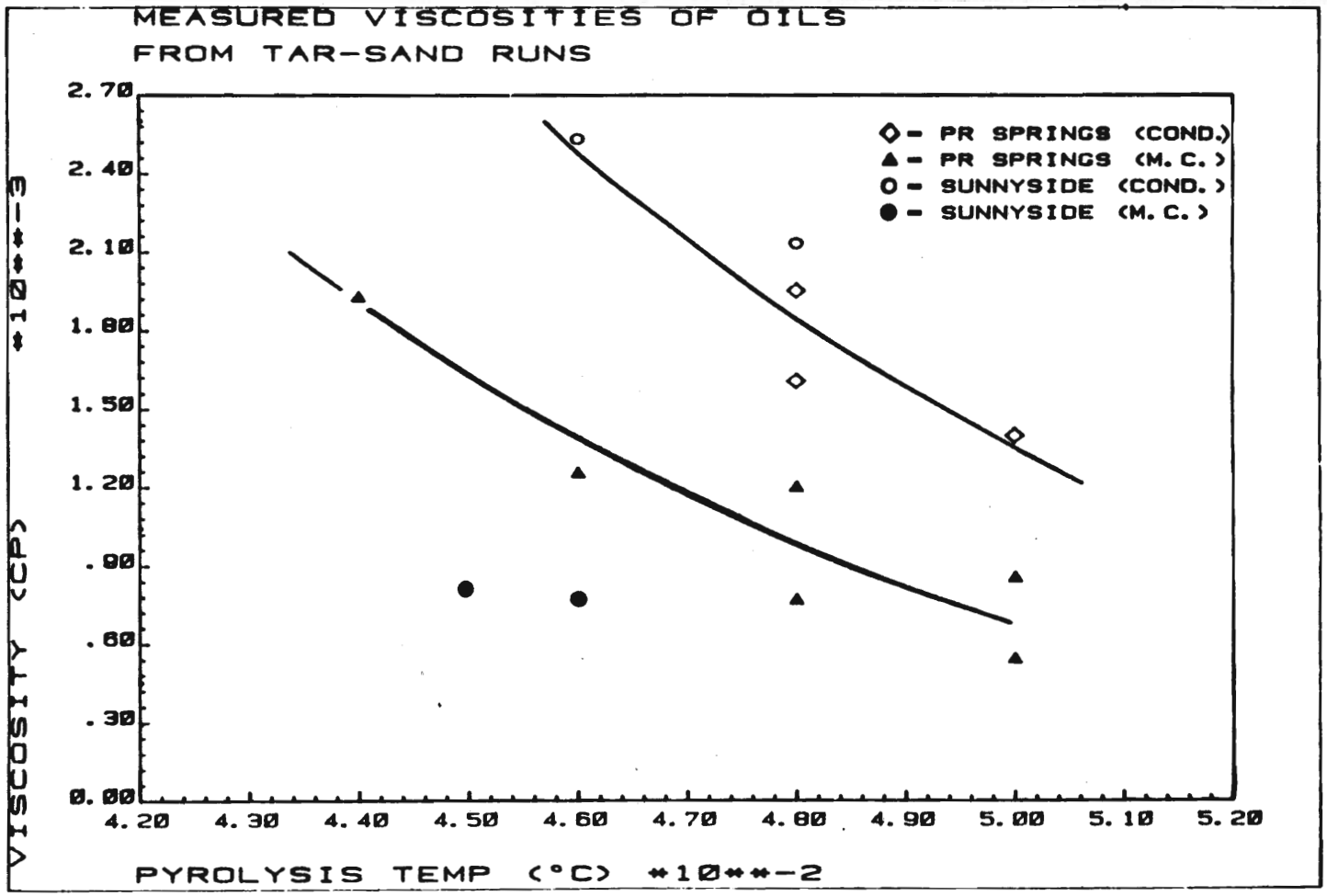


Figure 29. Measured viscosities of thermally recovered oils

Sulfur content varied from 0.19 to 0.24 weight % for the oils collected. Sulfur compounds are the predominant nonhydrocarbons in conventional crude stocks, and are undesirable because of corrosivity, chemical unstability, bed stench, and catlyst poisoning. Approximately half of the conventional crude oils, explored up to this point, have sulfur contents under 0.5% by weight (Neumann et al., 1981). Therefore, the sulfur contents for the products in this research are in the same general range as conventional crudes. There were no significant trends relating product sulfur content to pyrolysis-bed temperature. The data for sulfur analysis are found in Table 9 of Appendix A.

Nitrogen content was determined with the same instrumentation that determined carbon and hydrogen content; however, the nitrogen results tend to be much less accurate because the standard used had a much higher nitrogen content than that of the product samples. The nitrogen contents did stay near 0.7% by weight. The data for nitrogen analysis are in Table 9 of Appendix A.

Figure 30 shows the results of the simulated distillations of the product oils. Data from these runs are found in Table 11 of Appendix A. The simulated distillation for a sample of condenser oil, from run number 34, which was hydrotreated over a commercial catalyst at 430°C and 2300 psig, is also shown.

For the hydrotreatment at 430°C and 2300 psig, the oil sample was allowed to react for 4.5 hours. These conditions were similar to those used by Brechtel (1981). After the reaction, it was determined that 34% by weight of the original sample had been converted to gas. The remaining liquid experienced an increase in the C/H ratio from 0.63 to

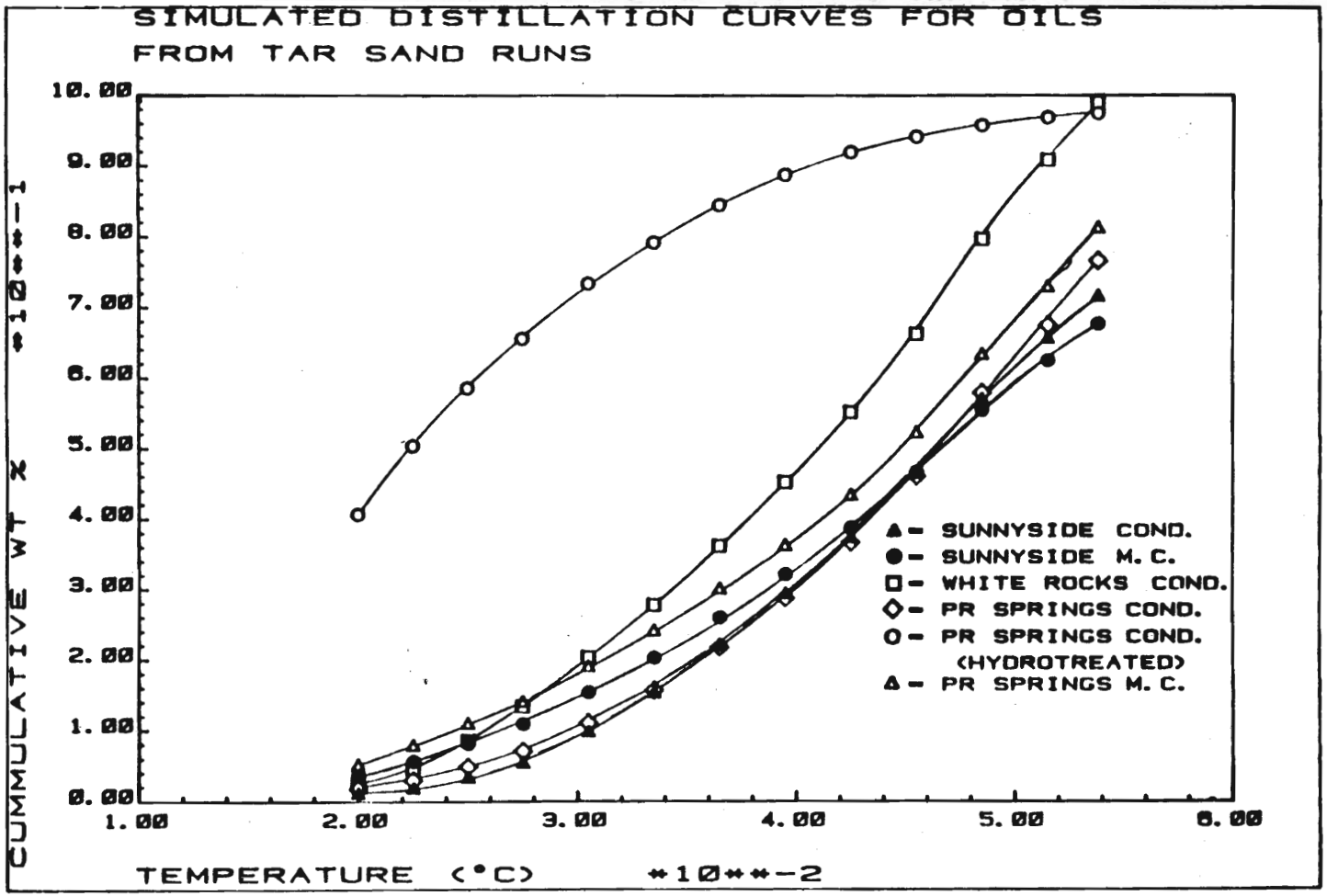


Figure 30. Simulated distillation curves of thermally recovered oils

0.82, and a decrease in viscosity, at 20°C, from 2000 to 2 cp. The liquid also decreased in density from 19 to 31.53 °API. Desulfurization was also apparent by a decrease in sulfur content from 0.165 to 0.02 weight percent. Deposition of coke onto the catalyst accounted for 6%, by weight, of the original sample. From these results, it appears that the hydrotreating run was done under conditions that were too severe, causing the sample to be cracked into lighter components. Further tests are needed to determine optimum hydrotreating conditions.

Energy Usage

A sample energy balance for run number 40 is shown in Appendix B. The heat, added to the reactor during the run, was measured experimentally, as was the heat loss to the surroundings. The largest source of error is assumed to be the amount of heat loss by conduction through the reactor walls. Figures 31 and 32 show the temperature profiles of the combustion and pyrolysis, respectively, showing that the various temperatures are maintained at a fairly constant value. From the results in Fig. 33 of Appendix B, 91% of the energy used was accounted for. It is also shown that each heat pipe transferred heat at a rate of 1400 Btu/hr, which is 35% of design heat transfer capacity. It can also be concluded that the heat pipes are capable of transferring 100% of the energy needed to maintain the upper bed at the desired temperature of pyrolysis. Also, for this run, there is more energy being lost to the surroundings, in the form of sensible heat and radial conduction, than electrical energy being added to the combustion bed to maintain process operation. This result agrees completely with the theoretical conclusions of Bezama (1983), who stated that the University of Utah

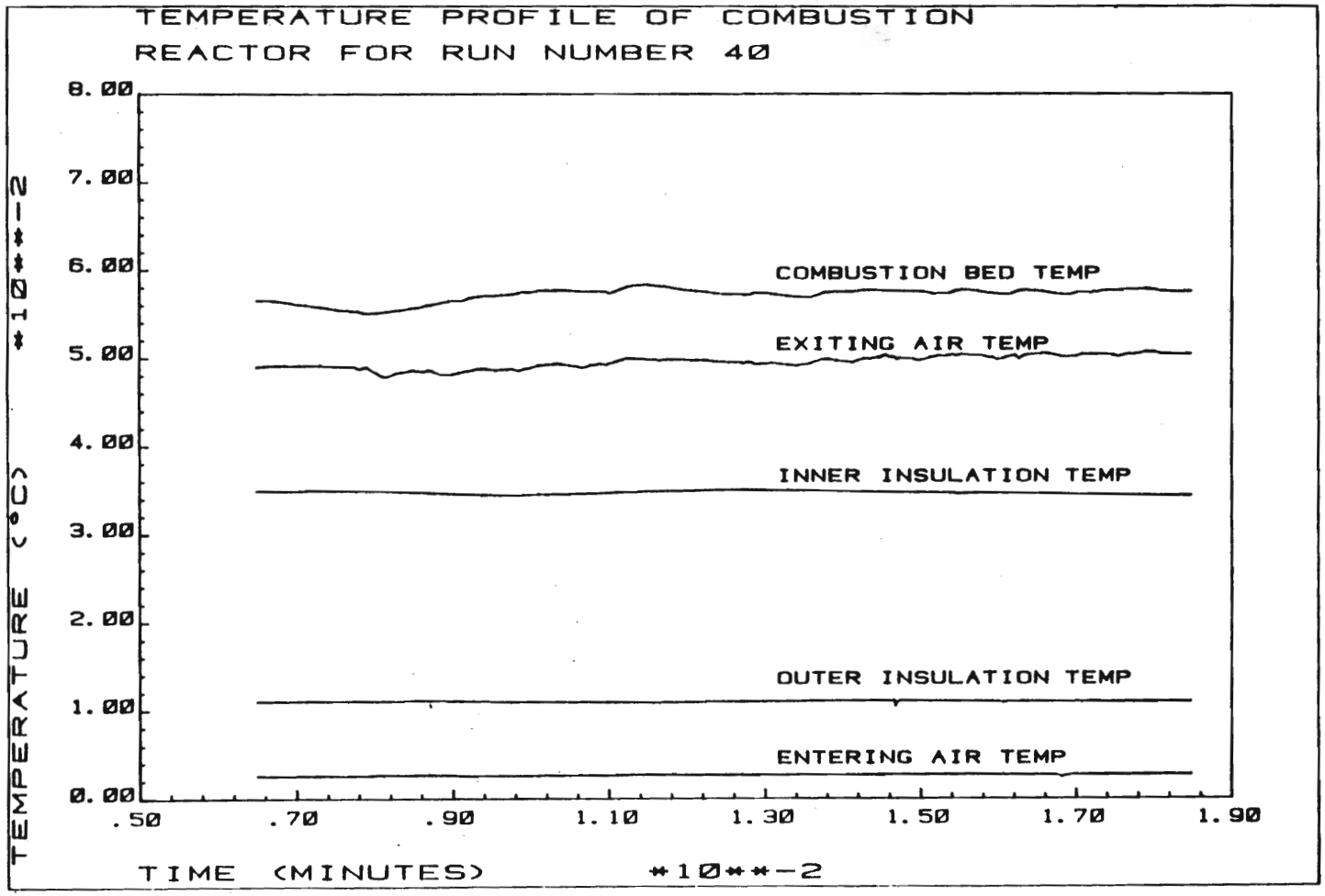


Figure 31. Temperature profile for combustion reactor from tar sand run recorded as FILE40

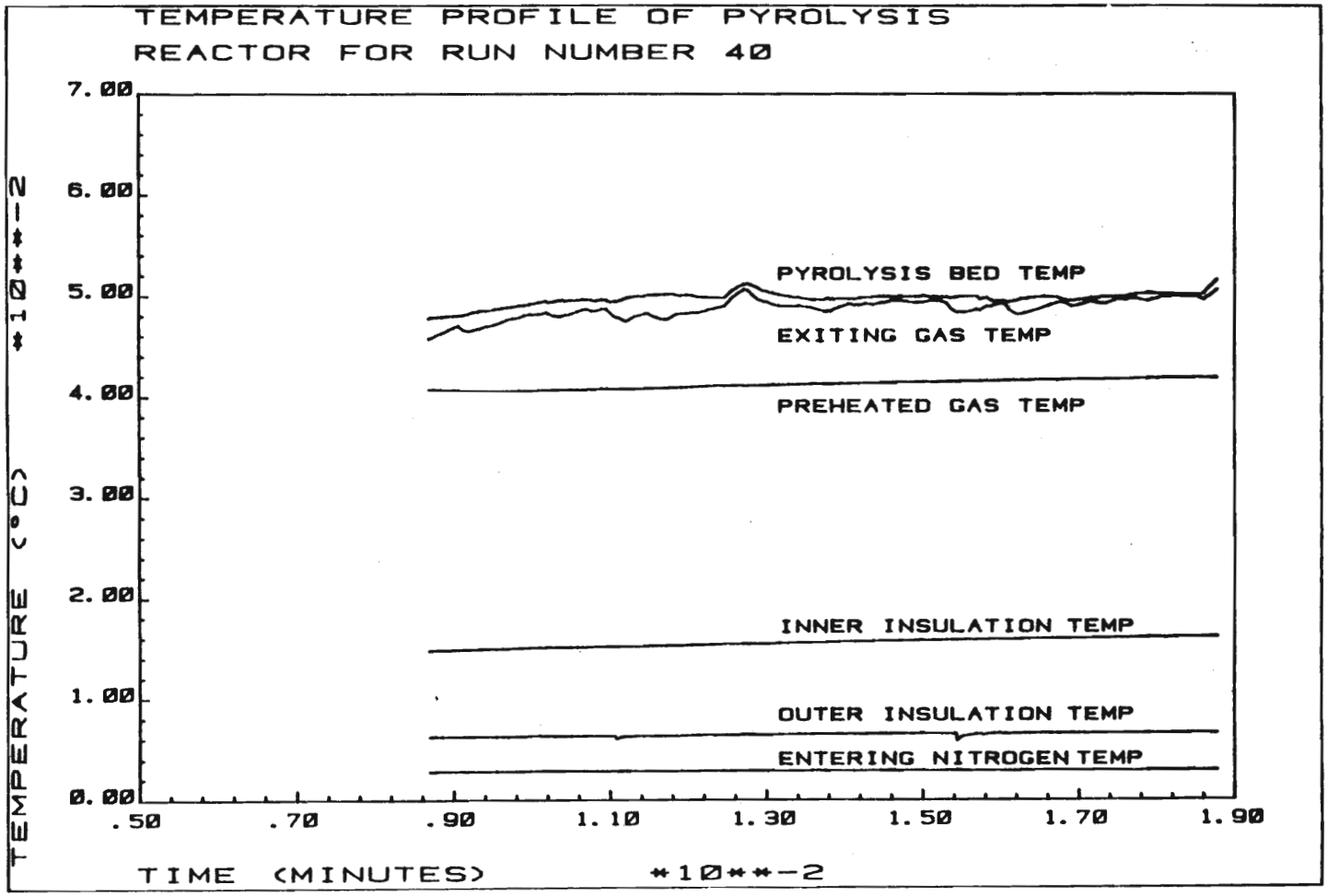


Figure 32. Temperature profile for pyrolysis reactor from tar sand run recorded as FILE40

Thermal Process could provide its own energy for operation, using tar sand with a minimum bitumen content of 8 weight percent, if the total lost work of the process is minimized. Furthermore, it can be assumed that, since each liquid-potassium heat pipe is engineered to transfer 4000 Btu/hr for these operating conditions, this experimental unit is capable of processing approximately 34 pounds of tar sand per hour, assuming that a similar bitumen content is used, and that the solids-transport systems can handle the increased flow.

CHAPTER VI

CONCLUSIONS AND RECOMMENDATIONS

A scaled-up replacement of a two-staged fluidized-bed reactor, for the thermal processing of tar sands, has been constructed, tested, and successfully operated over a range of operating conditions. The 4.25-inch-diameter reactors contain 3 liquid-metal heat pipes, with potassium as the working fluid, which function to transfer all of the energy needed to maintain the upper-bed temperature and to pyrolyze the bitumen. The conclusions made from the results of this research are listed below.

1. The three heat pipes provide a very effective means of heat transfer from the combustion bed to the pyrolysis bed. For the reactor and heat-pipe arrangement of this research, the system is capable of processing at least 30 pounds of tar sand per hour.

2. The University of Utah Thermal Process can be scaled-up without too much difficulty. An increase in reactor diameters, volumetric flows of the fluidizing gases, and number of heat pipes, enlarges the compartmentalized sand beds for both reactors to permit an increased processing capacity.

3. The temperatures vary by no more than 5°C within the fluidized bed, demonstrating uniformity of conditions for the fluidized beds.

4. Process control is improved by reactor scale-up. Digital PID control provides adequate control for both fluidized-bed level and

pyrolysis-bed temperature. Large solids flow improves operability of the solids-control valves and results in less plugging of the solids downcomers in the reactors. With the proper choice of valves and flow schemes, further scale-up should not inhibit process controllability.

5. There is enough energy, available from the heat of coke combustion, to maintain the bed temperatures and to pyrolyze the bitumen. However, the tar sand must contain at least 8 weight percent bitumen and sufficient energy must be recovered from the exiting spent sand, vapors, and gases. Furthermore, energy loss by heat transfer must be reduced. Further scale-up should facilitate more effective means for energy recovery.

6. This process is flexible with respect to different tar sands. The major difficulties encountered by using various tar sands were the behavior in the small screw feeder and the degree of fines accumulation on the solids-separation filter.

7. Liquid-crude yields as high as 71 weight percent, based on the original combustible bitumen of the tar sand, can be obtained from this process. The most important variable, which appears to affect product yield and quality, is the temperature of the pyrolysis reactor.

8. The products obtained from this process are significantly more amenable to handling and further upgrading than the original bitumen. The high temperature of the process greatly improves viscosity, API gravity, and boiling-point fractions.

Several recommendations can be given as a result of this work.

1. The operability of the present process could be further improved by the proper installation of a cyclone. This would reduce the

amount of solids loading on the sintered-metal filter before serious coke build up could occur. Positioning of the cyclone inside the entrainment section of the pyrolysis reactor is suggested, in order to maintain a sufficient cyclone temperature to prevent plugging by condensed oil.

2. Performance of this process could be improved by the installation of a better feeding system. The correct type of feeder would need to be chosen to assure adequate feeding capacity, feed-rate consistency, and flexibility with respect to different tar sands. Areas that could be investigated include use of a vibrating device, optimization of the angle of repose in the feed bin, or employment of a mechanical mixing device in the hopper. For feeds with higher agglomeration tendencies, other feeding methods besides the conventional screw-type feeders may need to be investigated.

3. The present system is amenable to further process-variable study. Further studies could determine the effect of fluidizing-gas flow rate, fluidizing gas composition, and average particle size of the tar-sand feed. A kinetic study of the pyrolysis of tar-sand bitumen is also recommended. A suitable model for the kinetics involved would aid process optimization.

4. Much longer runs should be made in order to study the process in the absence of start-up effects. Since fluidized beds closely model ideal back-mix reactors [Kunii et al., 1969)], a run duration of at least five average solids-residence times is required before the clean sand, that was originally in the pyrolysis reactor, is essentially

replaced with coked sand. After that time is elapsed, the process can be assumed to be operating at a steady state.

5. More study is recommended to determine the effects of erosion on the processing equipment. It is essential to know what corrosion rates can be expected for the outer heat-pipe walls. Application of abrasion-resistant coatings that can also withstand high fluidized-bed temperatures should also be investigated.

6. Different methods of heat addition to the combustion bed should be investigated. Electrical heating is, by far, the easiest way to add energy needed to maintain operation of the laboratory process. However, energy usage of the laboratory process could be better studied if the heat were added to the inside of the combustion bed via some sort of immersed heating probe. This type of heating would result in a significant decrease of heat loss from the combustion section of the process. More practical methods of heat addition will need to be developed for a pilot-plant process, such as the regulated input of a combustible substance into the combustion bed. Further improvement of the insulation around the reactors to reduce heat loss is also suggested.

7. The most significant way to reduce the thermodynamic irreversibility or lost work of this process involves the recovery of heat from the exiting sand of the combustion reactor. This heat could be used to produce steam, to preheat incoming tar-sand feed, or to heat water. Larger pilot-plant equipment is needed to study heat recovery from high-temperature flowing sand, where uninterrupted sand flow of sufficient quantity would occur.

This process is a stable, easily controlled method for hydrocarbon recovery from tar sands and should be studied on a larger, pilot-plant scale with the use of automatic process control and data acquisition. Finally, the applicability of this process should not be limited to tar sands alone, but should be considered as a method for processing other hydrocarbon sources such as biomass, coal, oil shale, etc.

APPENDIX A

EXPERIMENTAL DATA

Table 6

Operating Conditons for Tar Sand Runs
(Sunnyside Tar Sand)

Disc File Number	16	18	19	20
Wt % bitumen in tar sand	10.6	10.6	10.6	10.6
Length of run (hrs)	5.83	2.12	3.12	2.37
Average sand feed rate (lb hr)	8.8	13.2	12.9	14.0
Feed mix (lb tar sand/lb sand)	1.25	2.00	1.77	1.71
Total quantity of feed (lb)	51.5	28.0	40.25	33.25
Average combustion bed hold-up (lbs)	16.6	16.6	16.6	16.6
Average pyrolysis bed hold-up (lbs)	8.3	8.3	5.7	8.3
Fluidizing air flow for combustion bed (SCFH)	70.0	59.8	70.0	70.0
Fludizing nitrogen flow for pyrolysis bed (SCFH)	36.3	36.3	36.3	36.3
Pyrolysis bed temperature (°C)	460.0	450.0	450.0	480.0
Oil yield (wt %)	45.8	45.2	57.2	58.4
Coke yield (wt %)	13.9	18.2	18.2	NA

Table 7

Operating Conditions for Tar Sand Runs
(Whiterock Tar Sand)

Disc File Number	22	23
Wt % Bitumen in tar sand	8.8	8.8
Length of run (hours)	2.35	1.87
Average sand feed rate (lb/hr)	1.87	2.53
Total quantity of feed (lb)	21.2	40.0
Average combustion bed hold-up (lbs)	16.6	16.6
Average pyrolysis bed hold-up (lbs)	8.3	8.3
Fluidizing air flow for combustion bed (SCFH)	59.8	59.8
Fluidizing nitrogen flow for pyrolysis bed (SCFH)	36.3	46.0
Pyrolysis bed temperature (°C)	450.0	480.0
Oil yield (wt %)	53.5	56.0
Coke yield (wt %)	NA	NA

Table 8

Operating Conditions for Tar Sand Runs
(PR Springs Tar Sand)

Disc File Number	30	32	33	34	35
Wt % bitumen in tar sand	10.1	10.1	10.1	10.1	10.1
Length of run (hours)	2.02	2.13	1.87	1.83	1.87
Average tar sand feed rate (lbs/hr)	9.8	9.4	10.7	10.9	10.7
Total quantity of feed	19.7	20.0	20.0	20.0	20.0
Average combustion bed hold-up (lbs)	16.6	16.6	16.6	16.6	16.6
Average pyrolysis bed hold-up (lbs)	8.3	8.3	8.3	8.3	8.3
Fluidizing air flow for combustion bed (SCFH)	95.8	70.0	70.0	70.0	70.0
Fludizing nitrogen flow for pyrolysis bed (SCFH)	46.0	46.0	46.0	46.0	46.0
Pyrolysis bed temperature (°C)	440.0	460.0	480.0	500.0	480.0
Total energy input to heaters in combustion bed (Btu x 10^{-4})	1.10	1.62	2.09	2.27	1.78
Total energy input to heaters in combusdtion bed (Btu x 10^{-4})	1.10	1.62	2.09	2.27	1.78
Oil yield (wt %)	60.2	67.3	66.8	71.0	67.8
Coke yield (wt %)	28.5	21.9	16.4	16.0	13.0
Gas yield (wt %)	--	4.7	7.3	3.2	5.8
% Accounting of bitumen	--	93.4	90.2	89.9	86.4

Table 8 (continued)

Operating Conditions for Tar Sand Runs
(PR Springs Tar Sand)

Disc File Number	36	39	40	41
Wt % bitumen in tar sand	10.1	10.1	10.1	10.1
Length of run (hours)	1.45	1.18	1.70	1.97
Average tar sand feed rate (lbs/hr)	13.3	16.9	11.8	10.2
Total quantity of feed (lbs)	19.25	20.0	20.0	20.0
Average combustion bed hold-up (lbs)	16.6	16.6	16.6	16.6
Average pyrolysis bed hold-up (lbs)	8.3	8.3	5.7	8.3
Fluidizing air flow for combustion bed (SCFH)	70.0	70.0	70.0	70.0
Fluidizing nitrogen flow for pyrolysis bed (SCFH)	46.0	46.0	46.0	46.0
Pyrolysis bed temperature (°C)	500.0	500.0	500.0	500.0
Total energy input to heaters in combustion bed (Btu x 10^{-4})	1.51	1.41	1.77	2.06
Oil yield (wt %)	69.4	63.4	68.3	70.3
Coke yield (wt %)	13.1	14.8	16.1	--
Gas yield (wt %)	6.9	9.9	6.2	10.2
% Accounting of bitumen	89.2	87.9	90.3	--

Table 9

Elemental Compositions of Tar Sand Products
(Weight Percent)

Run Number	Condenser Oil				Mist Collector Oil			
	C	H	N	S	C	H	N	S
Sunnyside								
16	87.4	11.3	0.44	0.19	90.0	11.6	0.43	0.23
17	--	--	--	0.26	86.8	11.3	0.45	0.22
18	86.5	11.0	0.38	0.25	--	--	--	--
19	87.1	11.2	0.45	0.22	86.7	11.4	0.29	--
Whiterocks								
22	87.1	10.5	0.99	0.21	--	--	--	--
23	86.1	11.0	0.86	0.19	84.2	11.0	0.64	0.21
Pr Spring								
30	87.3	11.5	1.44	0.22	87.5	11.8	1.28	0.21
32	87.8	11.5	0.96	0.20	87.4	11.4	0.86	0.21
33	87.8	11.4	1.02	0.18	87.4	11.4	0.80	0.20
34	87.6	11.6	0.71	0.17	87.7	11.5	0.72	0.15
35	87.5	11.6	0.75	0.20	87.5	11.6	0.80	0.16

Table 10

Viscosities and API Gravities of Tar Sand Products

Run Number	API Gravity Condenser	API Gravity Mist Collector	Viscosity Condensers cps (25°C)	Viscosity Mist Collector cps (25°C)
16	19.2	21.5	2534	774
17	17.1	21.3	1475	817
18	17.6	--	1554	--
19	17.8	17.9	--	517
20	18.1	--	2133	--
22	18.4	--	--	--
23	19.4	19.9	1187	329
30	19.4	20.7	--	1930
32	19.5	21.8	--	1256
33	19.65	20.6	1606	771
34	19.1	20.8	2020	855
35	19.6	20.1	1952	1202
36	17.3	18.8	1750	716
39	18.1	19.7	1354	394
40	18.0	18.9	1354	723
41	17.5	18.9	1395	544

Table 11

Simulated Distillations of Synthetic Crudes
Cumulative Weight Percent

Tar sand	Sunnyside		Whiterocks	PR Springs		
Run Number	16		22	34		
Sample*	Con.	M.C.	Con.	Con.	M.C.	Con.**
Cut point (°C)						
200	1.2	3.6	2.4	1.8	5.2	40.6
225	1.9	5.6	4.7	3.0	7.9	50.3
250	3.4	8.1	8.6	4.9	11.0	58.5
275	5.5	10.9	13.1	7.1	14.1	65.5
305	9.9	15.4	20.4	11.1	19.1	73.3
335	15.4	20.3	27.8	15.7	24.2	79.1
365	22.1	26.0	36.2	21.8	30.1	84.4
395	29.5	32.1	45.2	28.8	36.3	88.7
425	37.4	38.7	55.1	36.7	43.4	91.9
455	46.5	46.6	66.2	46.0	52.2	94.1
485	57.0	55.2	79.6	57.8	63.3	95.7
515	65.6	62.4	90.8	67.4	72.9	96.8
538	71.6	67.6	99.0	76.5	81.2	97.4

* Cond. represents oil obtained from condensers, while M.C. represents oil obtained from the mist collector.

** This sample was hydrotreated over Ni-Mo catalyst.

Table 12

Thermodynamic Data Base for Enthalpy Calculations

Variable	Value	Units	Reference
Cp, sand (25°-580°C)	0.43	Btu/lb-°C	Bezama (1983)
Cp, coked sand (25°-500°C)	0.43	Btu/lb-°C	Bezama (1983)
Cp, coke (25°-500°C)	0.57	Btu/lb-°C	Bezama (1983)
Cp, bitumen (25°-500°C)	1.08	Btu/lb-°C	Bezama (1983)
Cp, oil (25°-500°C)	1.07	Btu/lb-°C	Bezama (1983)
Cp, gas (25°-500°C)	1.23	Btu/lb-°C	Bezama (1983)
Cp, nitrogen (25°-495°C)	1.23	Btu/gmol-°C	Bezama (1983)
Cp, air (25°-495°C)	0.0296	Btu/gmol-°C	Felder, et al. (1978)
Cp, CO2 (25°-575°C)	0.40	Btu/gmol-°C	Felder, et al. (1978)
Cp, O2 (25°-575°C)	0.029	Btu/gmol-°C	Felder, et al. (1978)
ΔH pyrolysis (25°C) PR Spring bitumen	200	Btu/lb	Bezama (1983)
ΔH combustion (25°C) PR Spring coke	14720	Btu/lb	Bunger (1979)
Thermal conductivity, Calcium silicate (80°-160°C)	0.0734	Btu/hr-ft-°C	Pabco Insulation
(110°-350°C)	0.0866	Btu/hr-ft-°C	Pabco Insulation

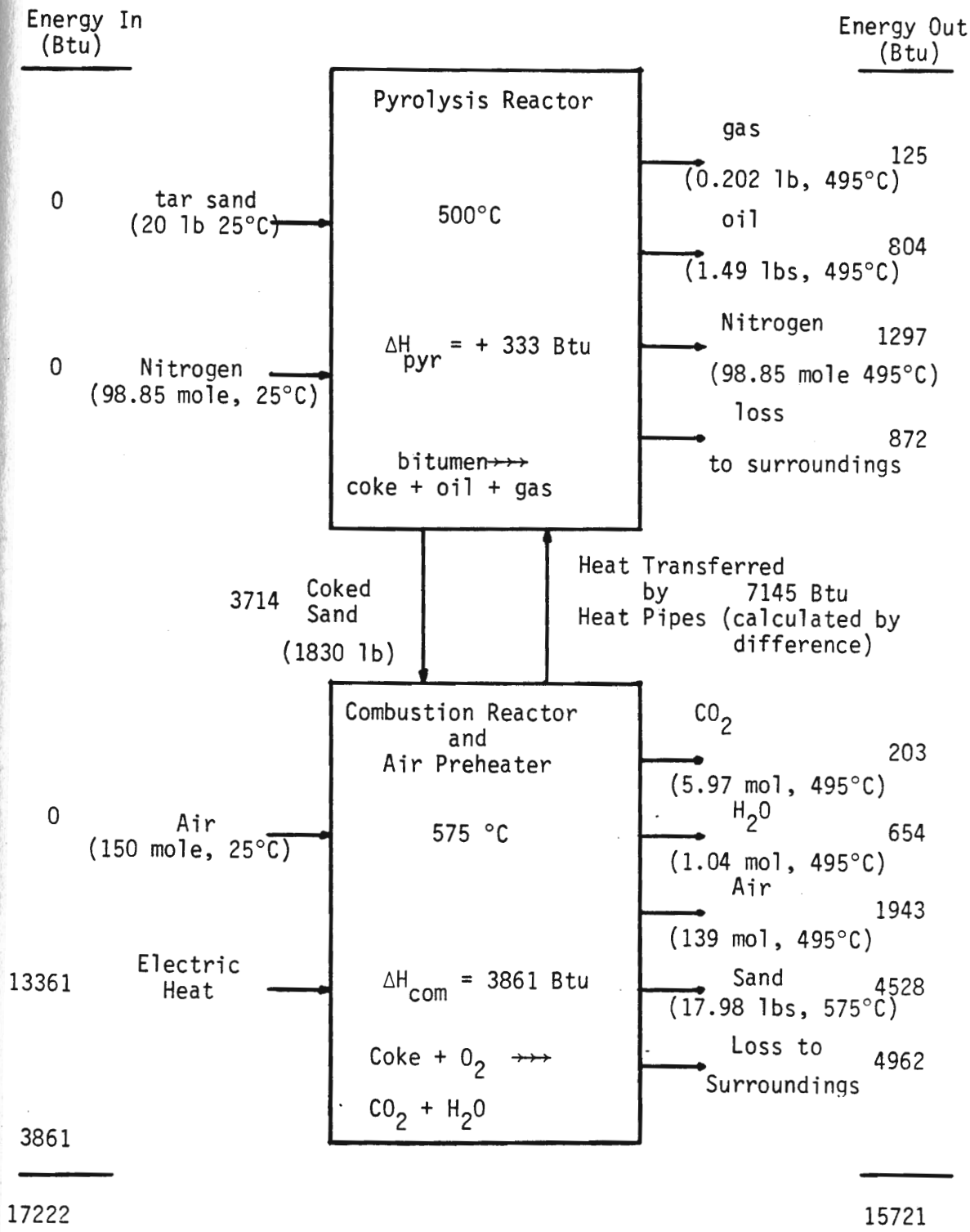


Figure 33. Material and enthalpy balance for tar sand run recorded as FILE40

APPENDIX B

EXPERIMENTAL ENTHALPY BALANCE

An attempt was made in this research to experimentally measure the heat usage for the tar sand runs. The following assumptions and specifications were used in making the calculations used for the sample enthalpy balance in this section.

1. The system is defined as the pyrolysis reactor, the intermediate zone between the upper and lower beds, the combustion reactor, and the air preheater.

2. The system operates at steady state for the entire length of the run.

3. The exit temperatures of the coked sand and spent sand are the same as their respective bed temperatures.

4. A negligible amount of heat exits the top or bottom of the system via conduction through the walls.

5. Coke is completely combusted in the lower bed.

6. Heat loss ($Btu/hr/ft^2$) is uniform for the outer insulated surfaces of the combustion reactor, intermediate zone, and air preheater. Similarly, heat loss is uniform for the outer insulated surface of the pyrolysis reactor.

7. This calculation is based on data from run #40, which processed twenty pounds of PR Spring tar sand (10.1 wt% bitumen).

8. Bitumen is assumed to be converted to 74% oil, 16% coke, and 10% gas.

9. Current and potential is constant for the outer heating wire of the combustion reactor. Negligible energy loss is assumed through the SCR power controller for the inner heating wire of the combustion

reactor. Potential and resistance is assumed constant for the inner heating wire of the combustion reactor.

10. The thermodynamic data base for these calculations is shown in Table 12.

APPENDIX C

HISTORIES OF TAR SAND RUNS

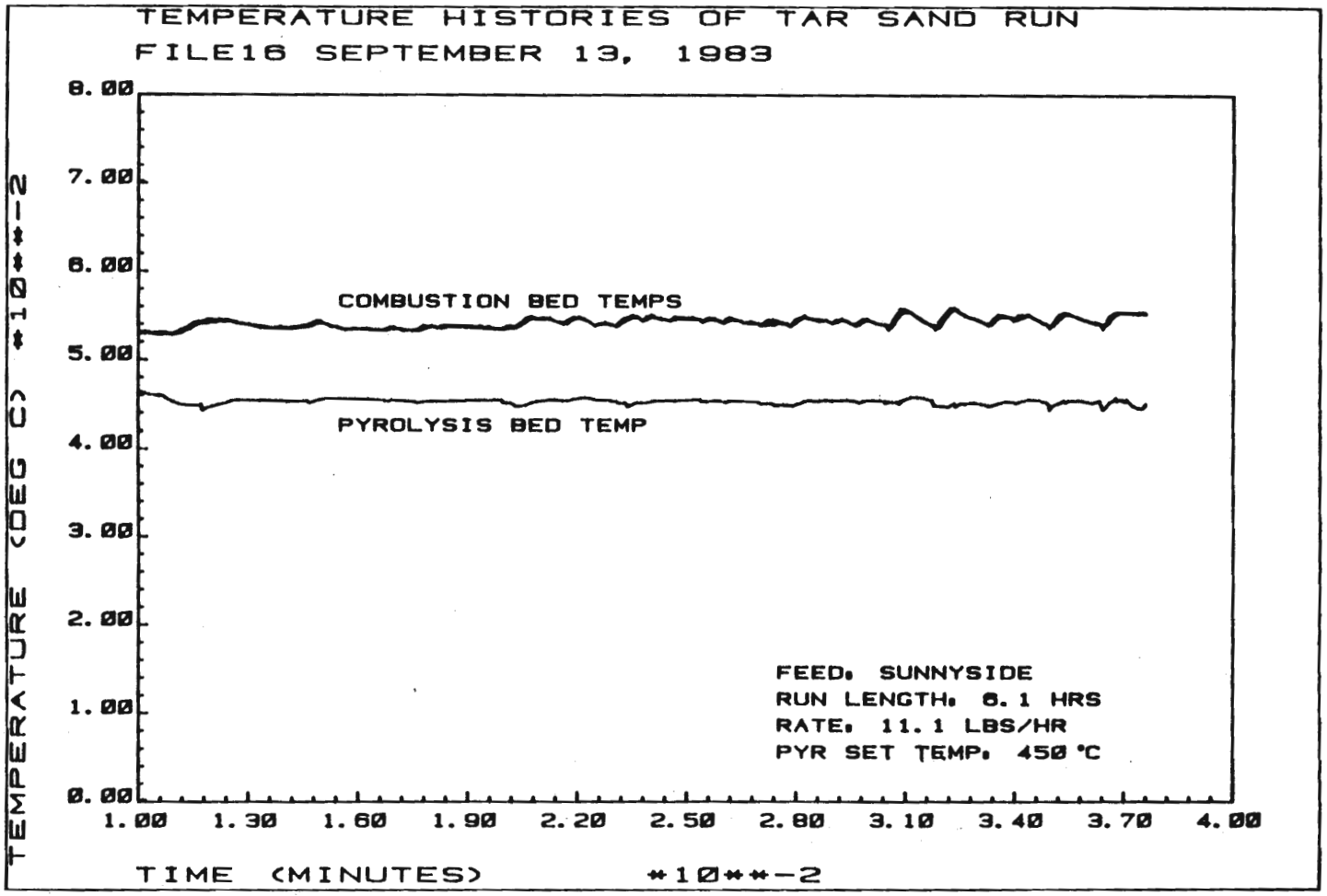


Figure 34. Temperature and bed level histories of tar sand runs

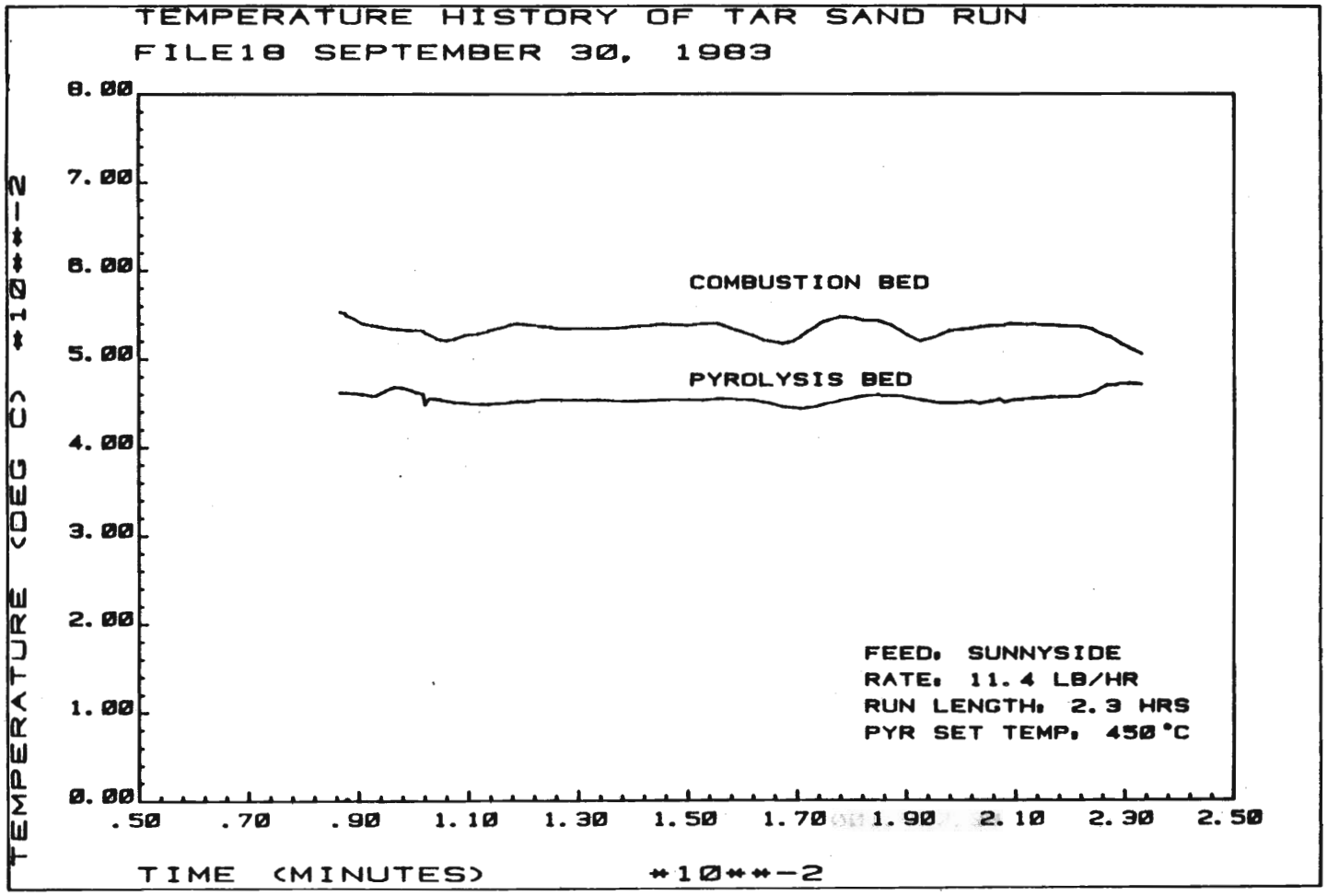
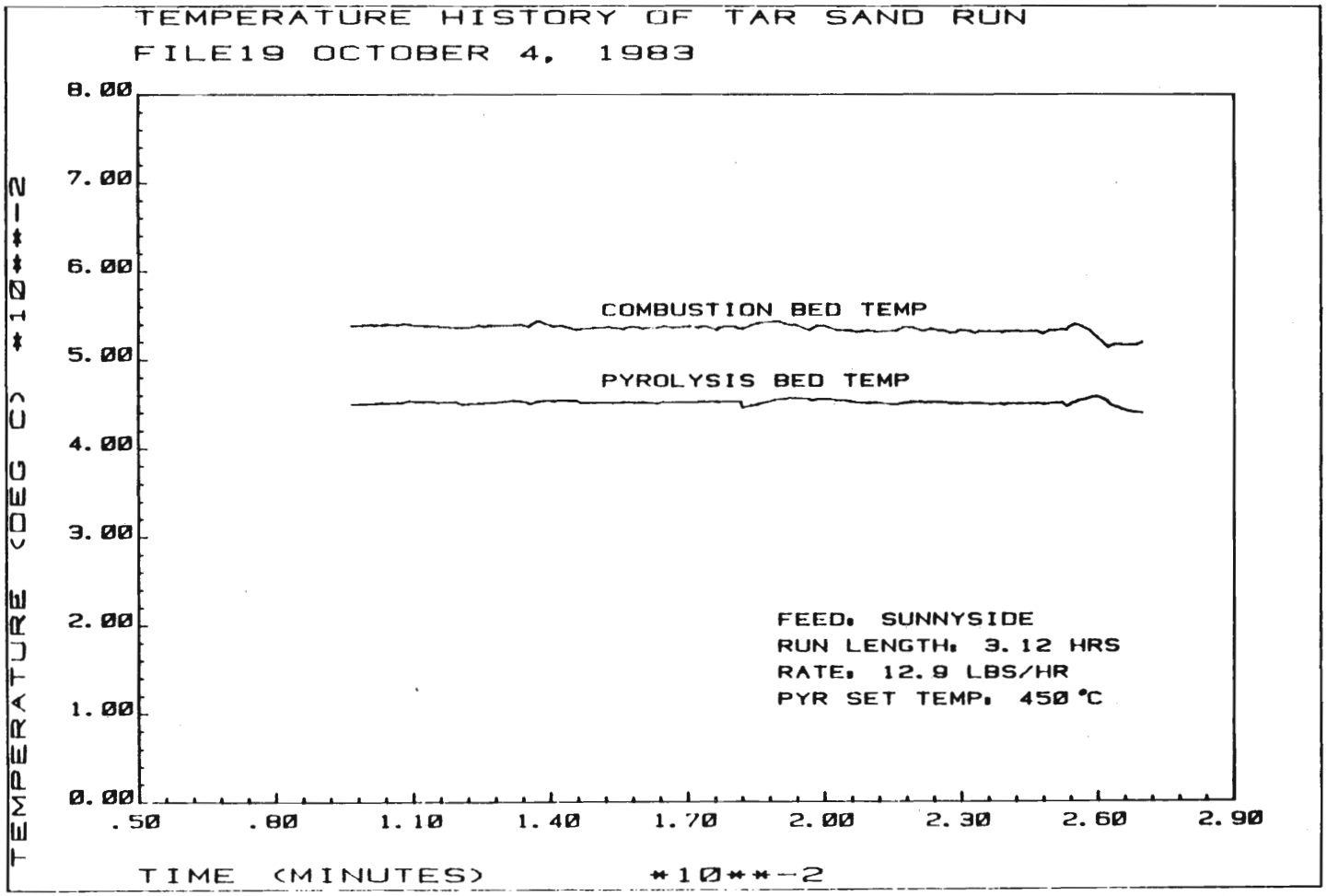


Figure 34. (continued)



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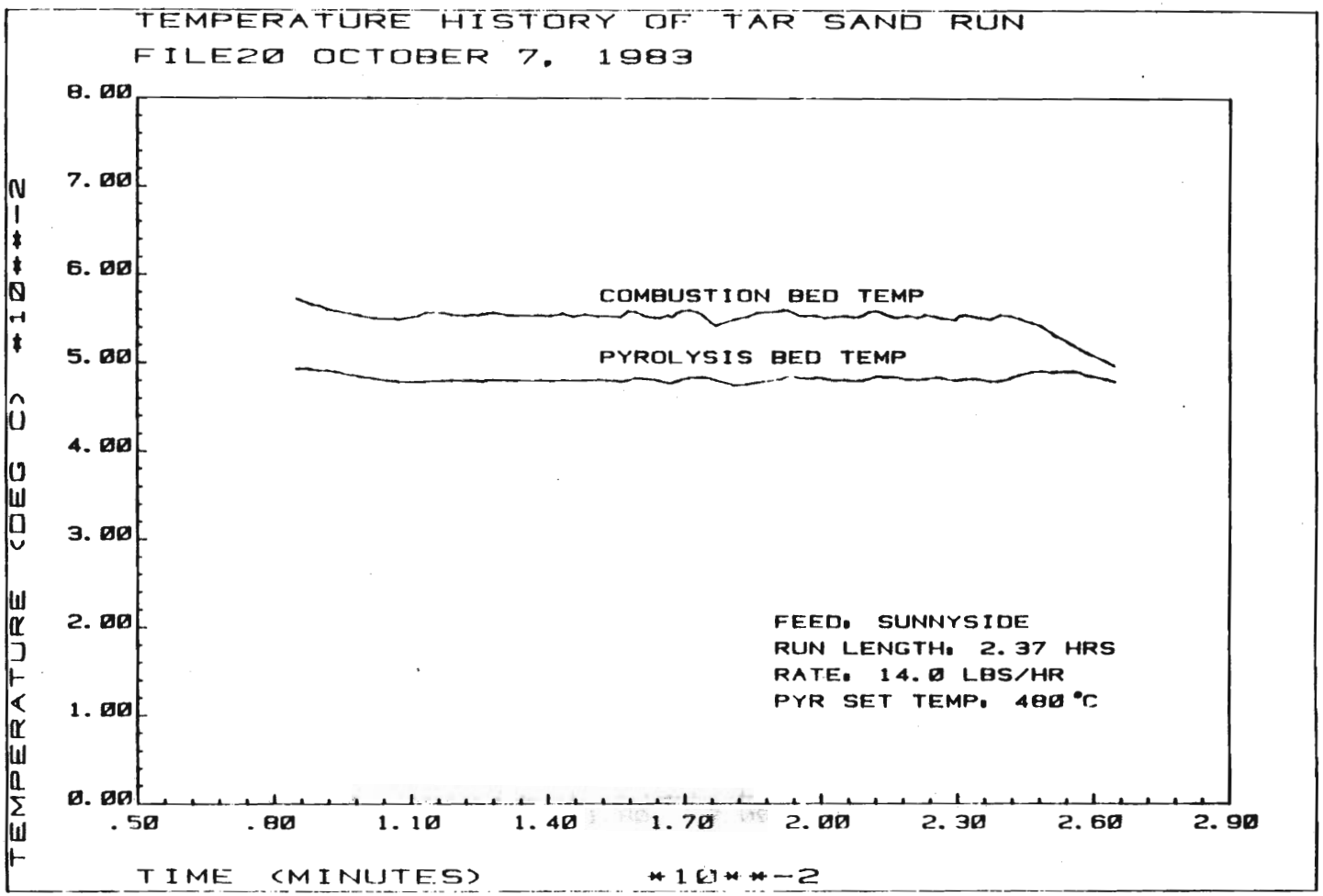

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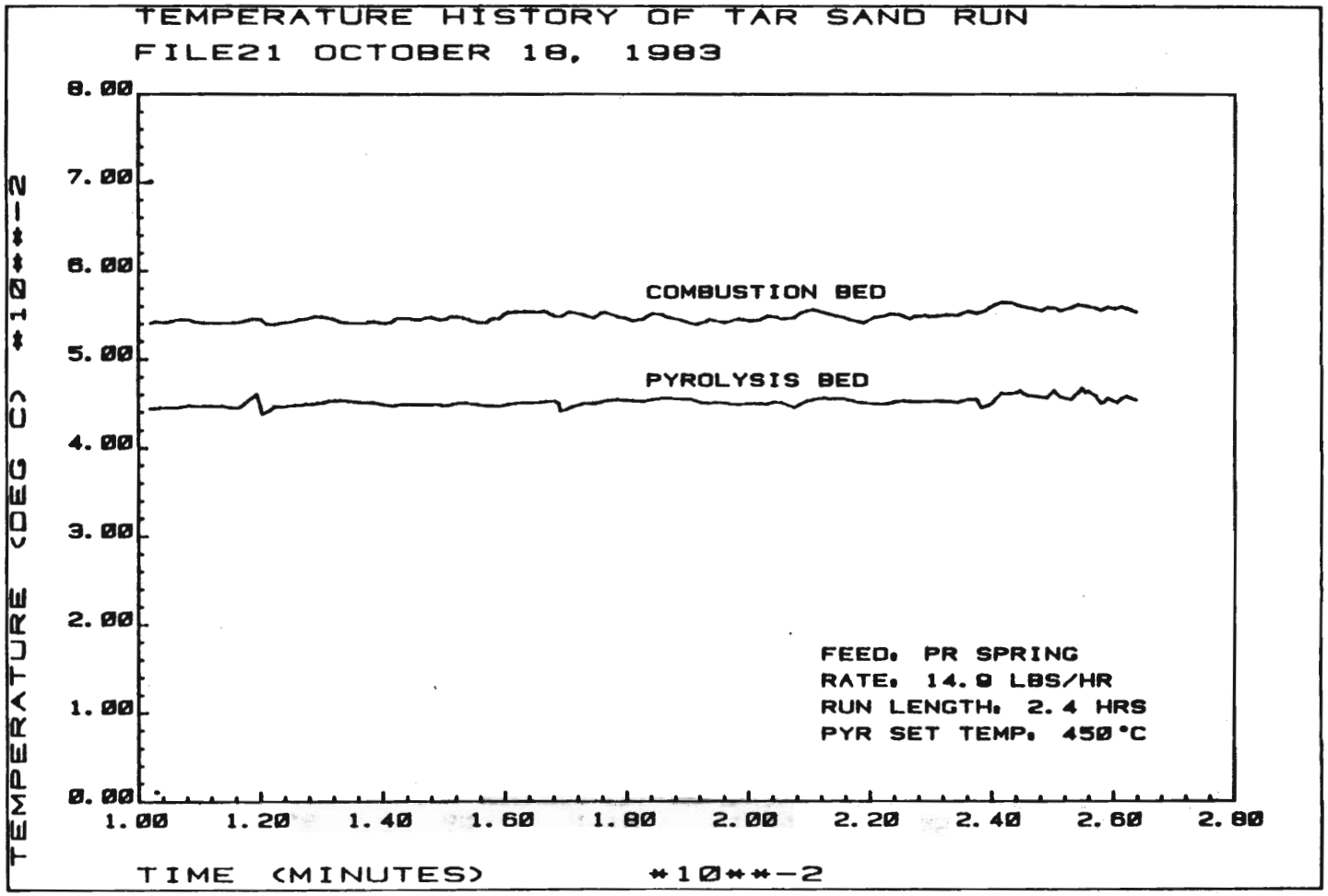

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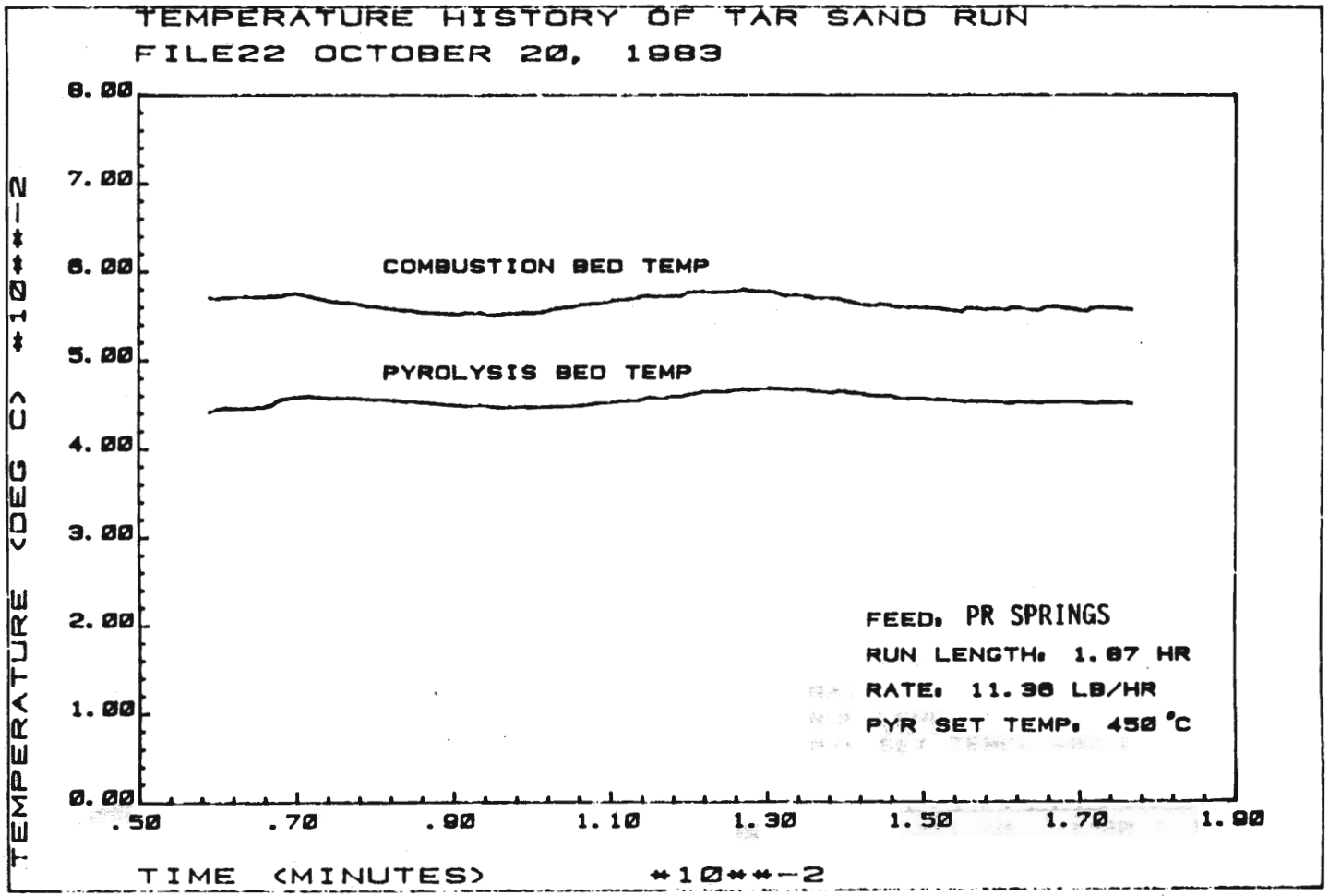

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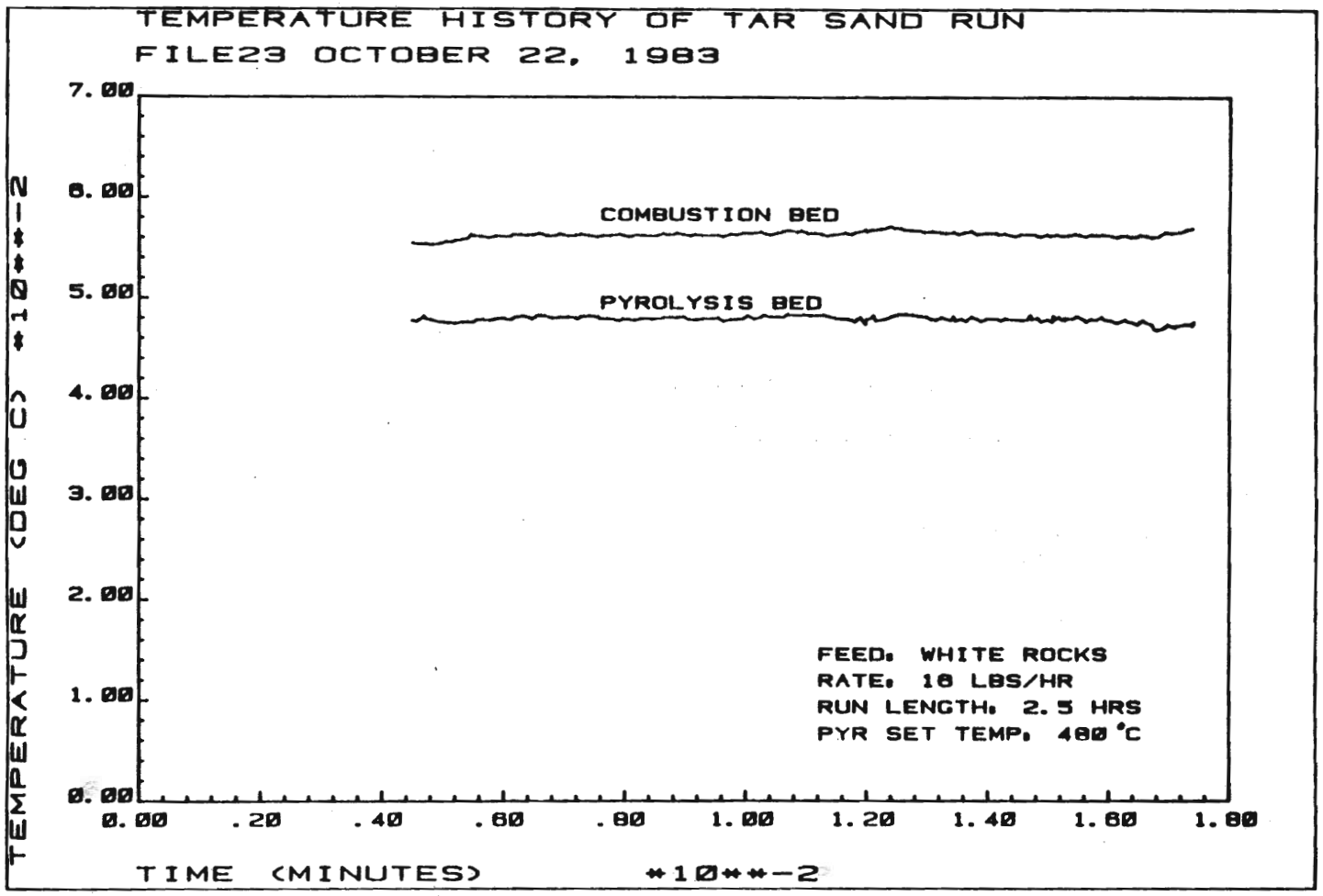

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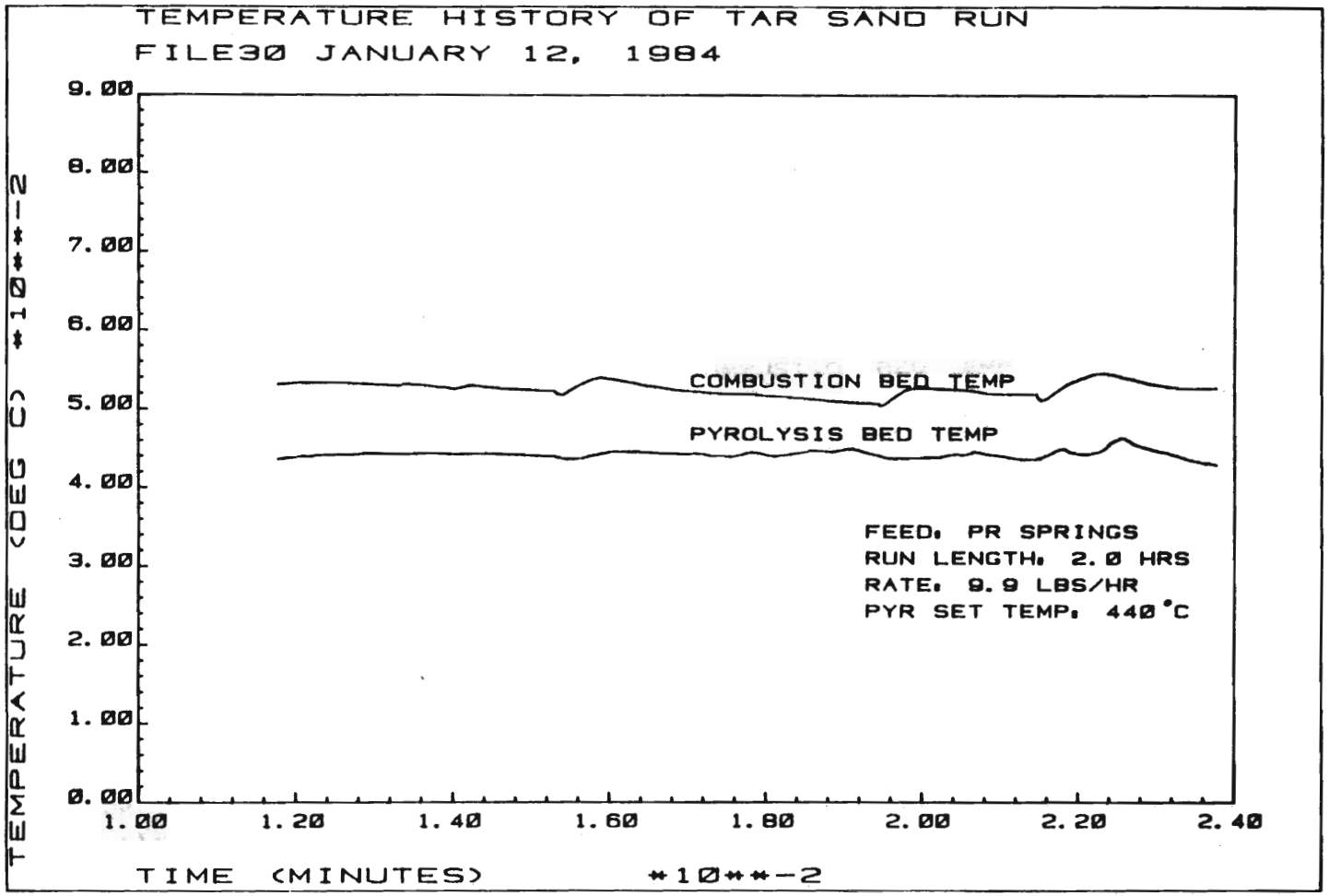

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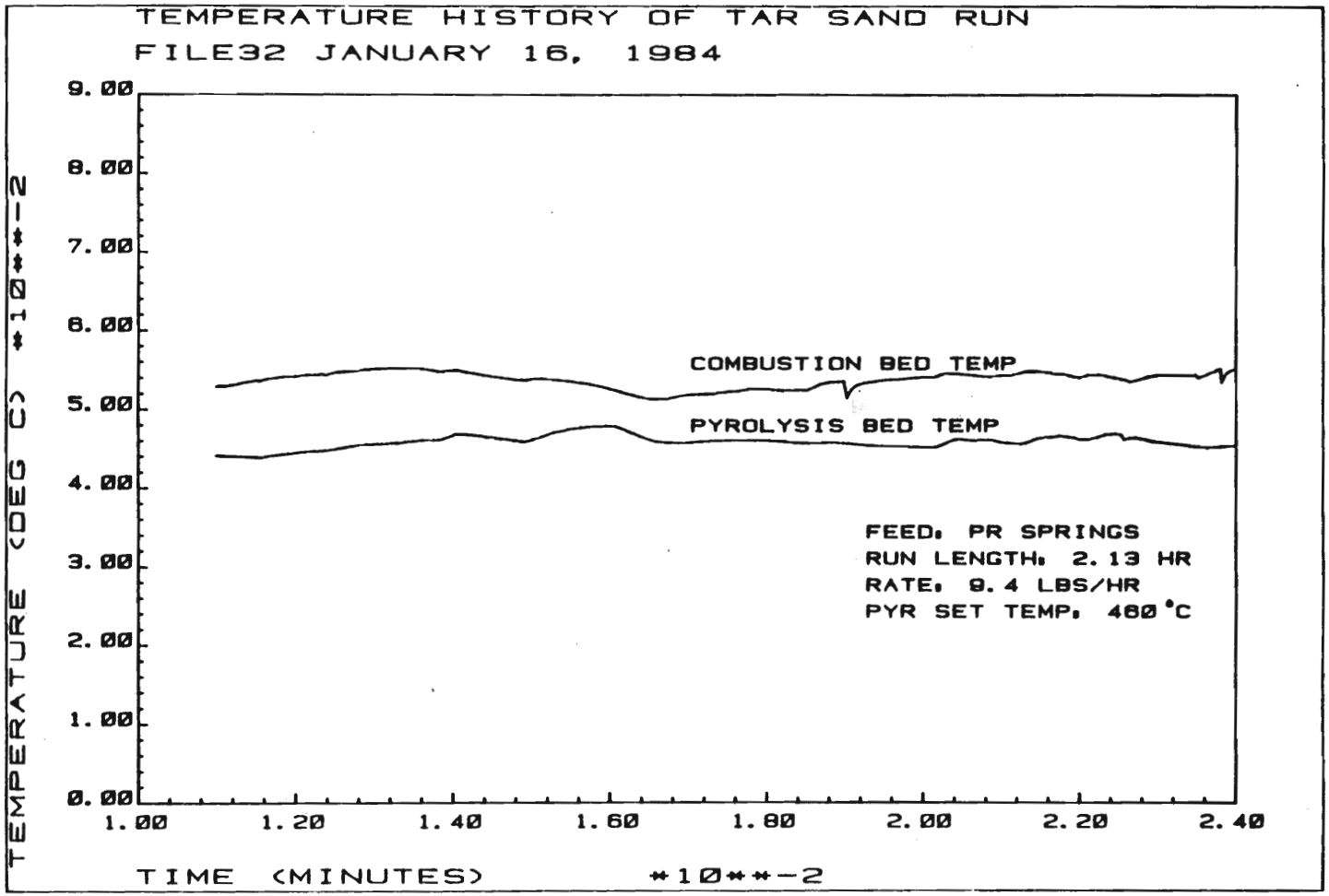

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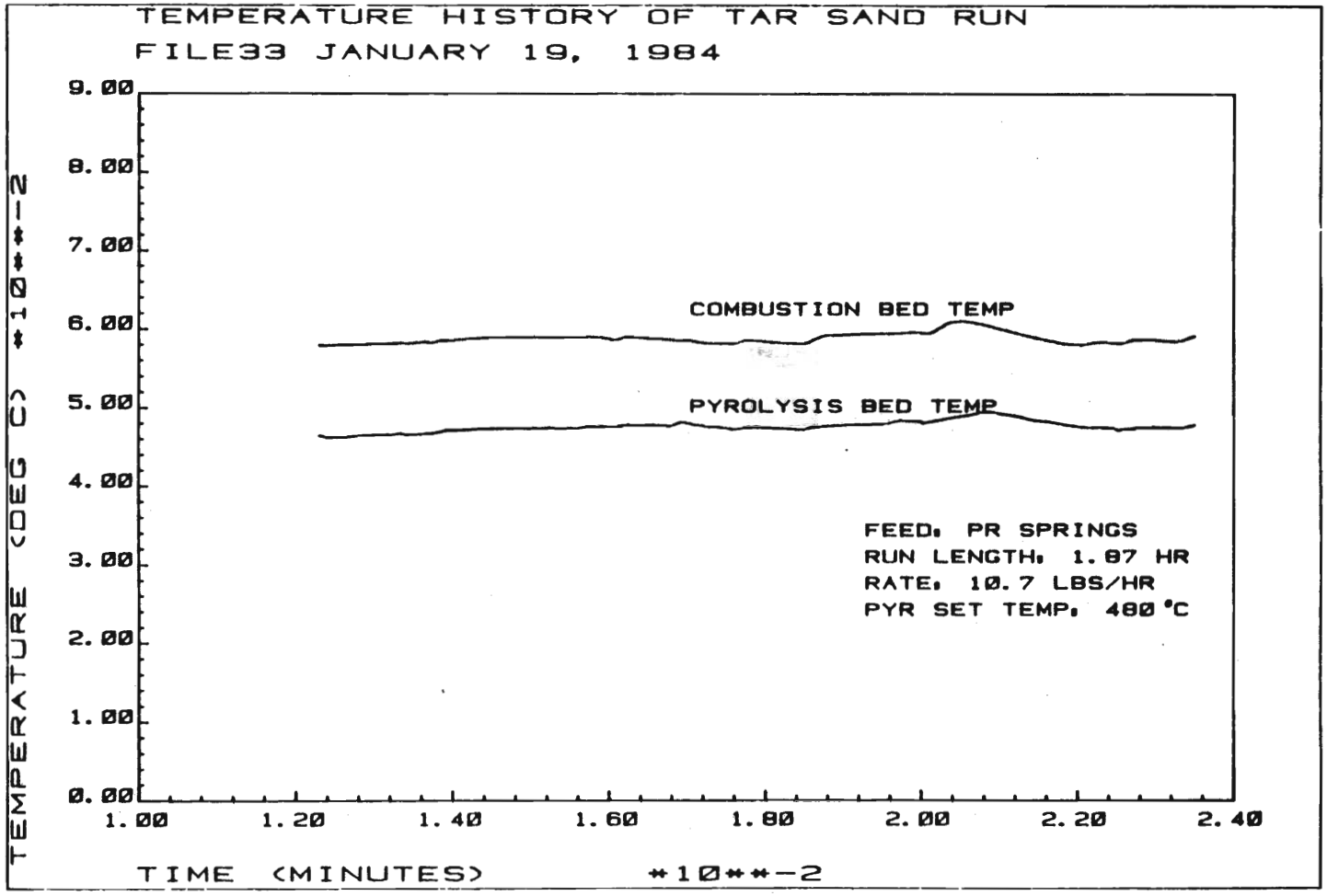

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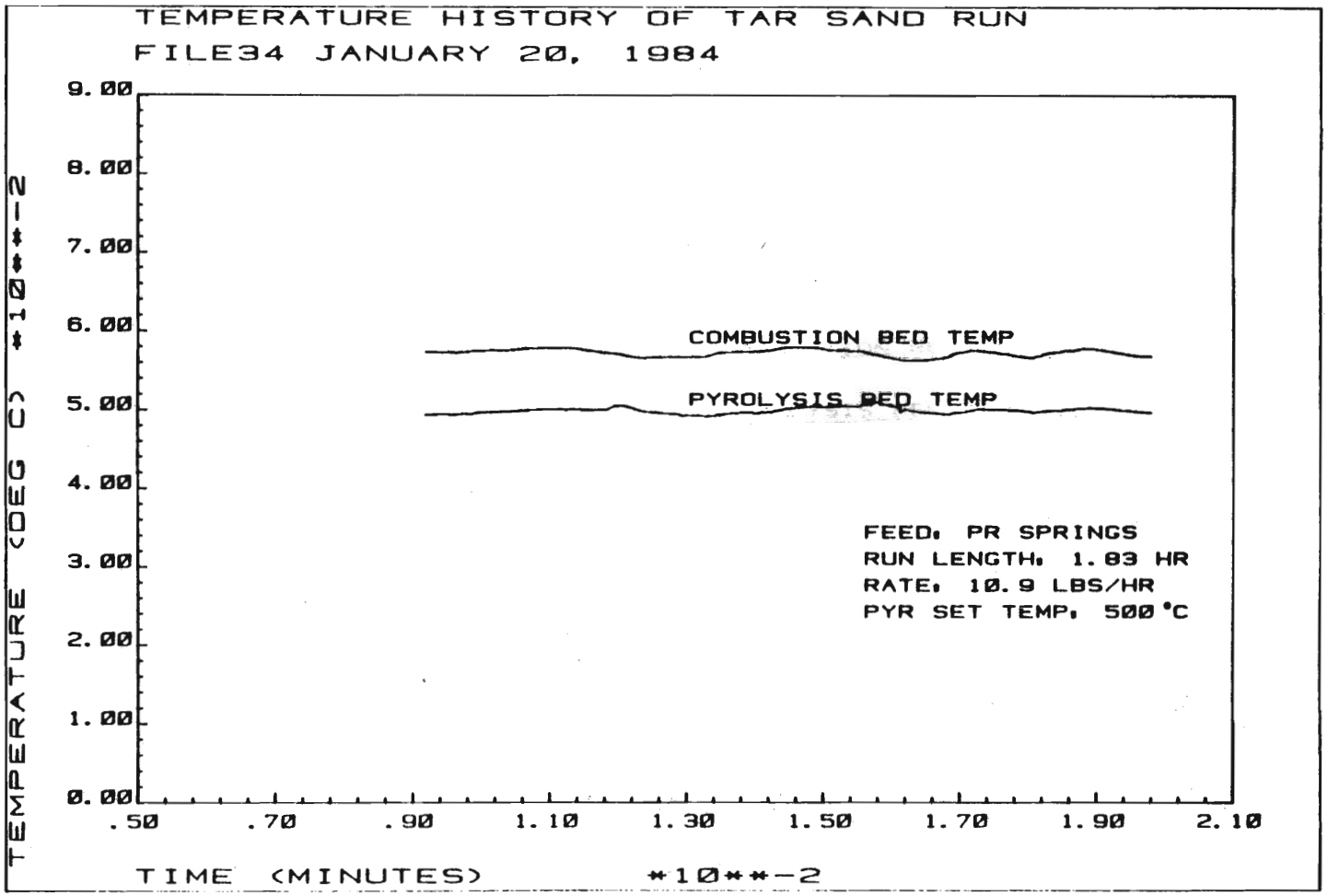

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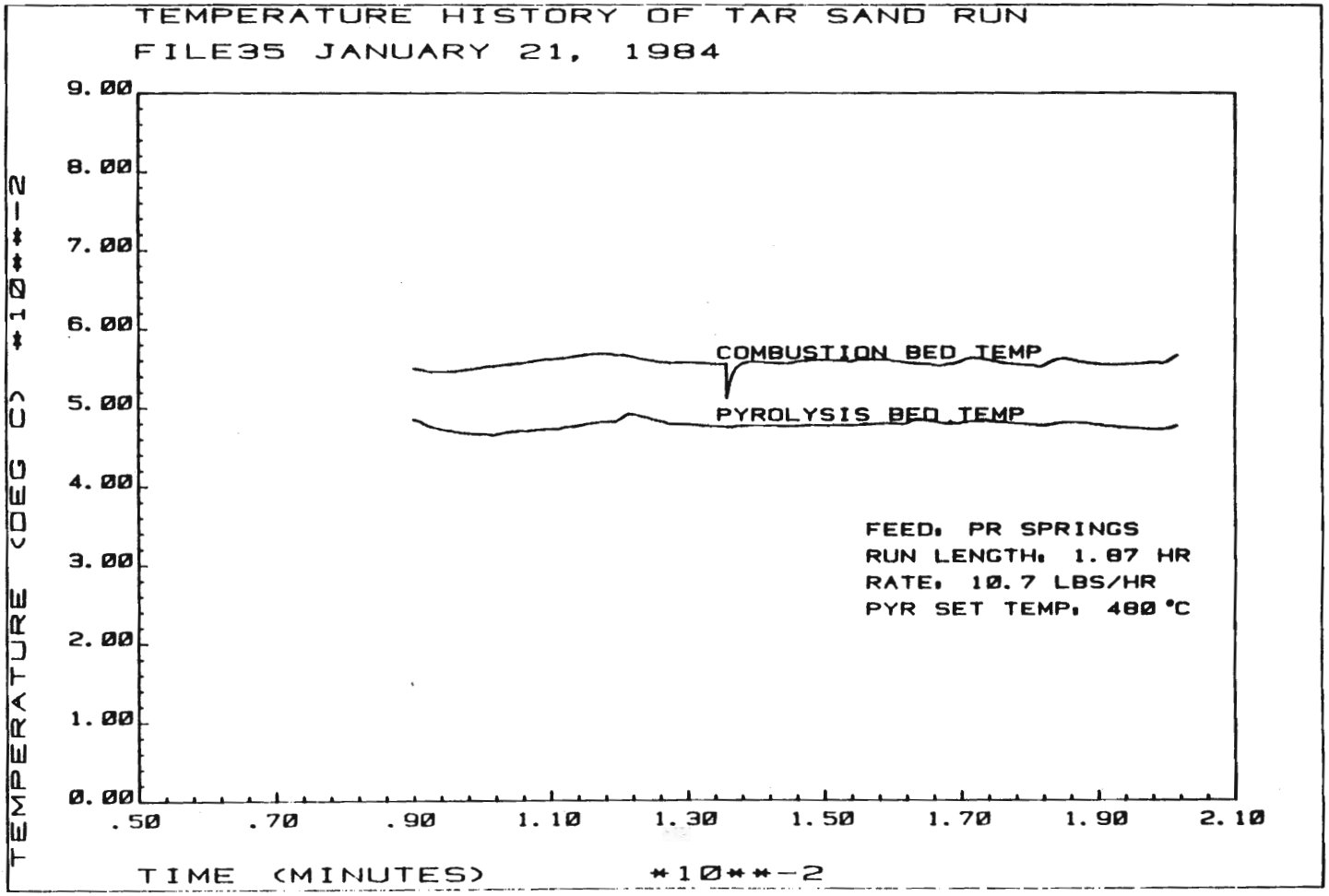

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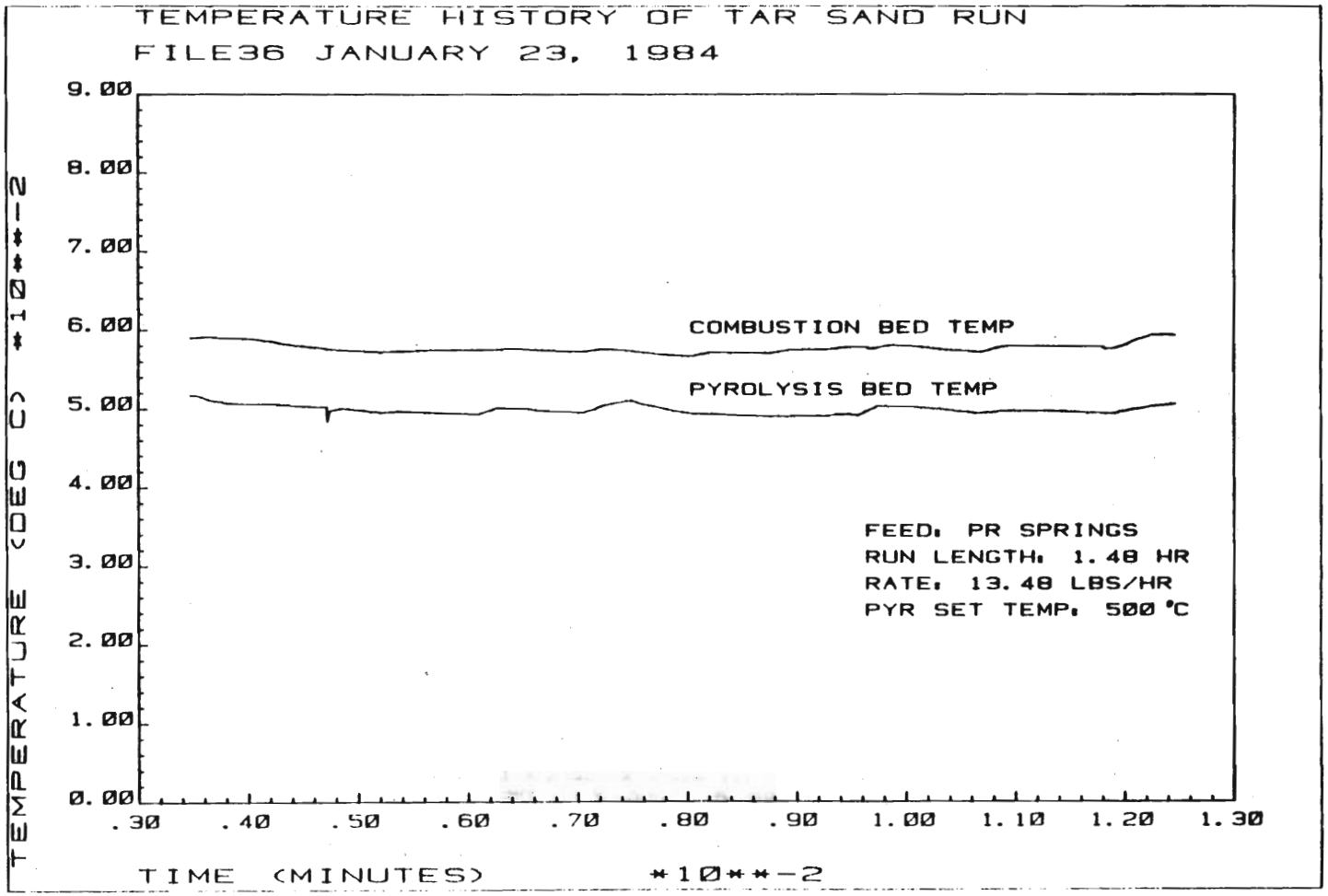

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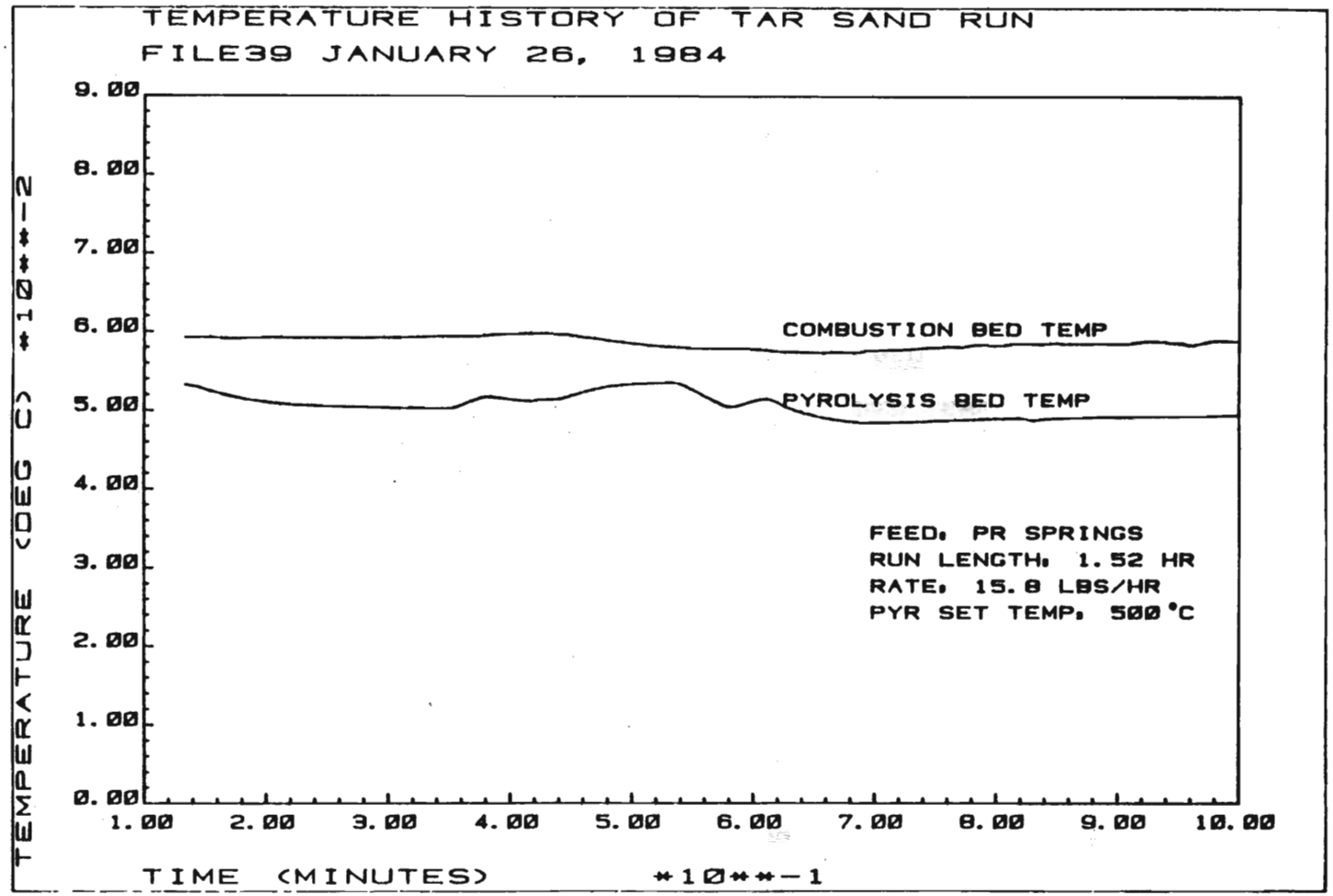

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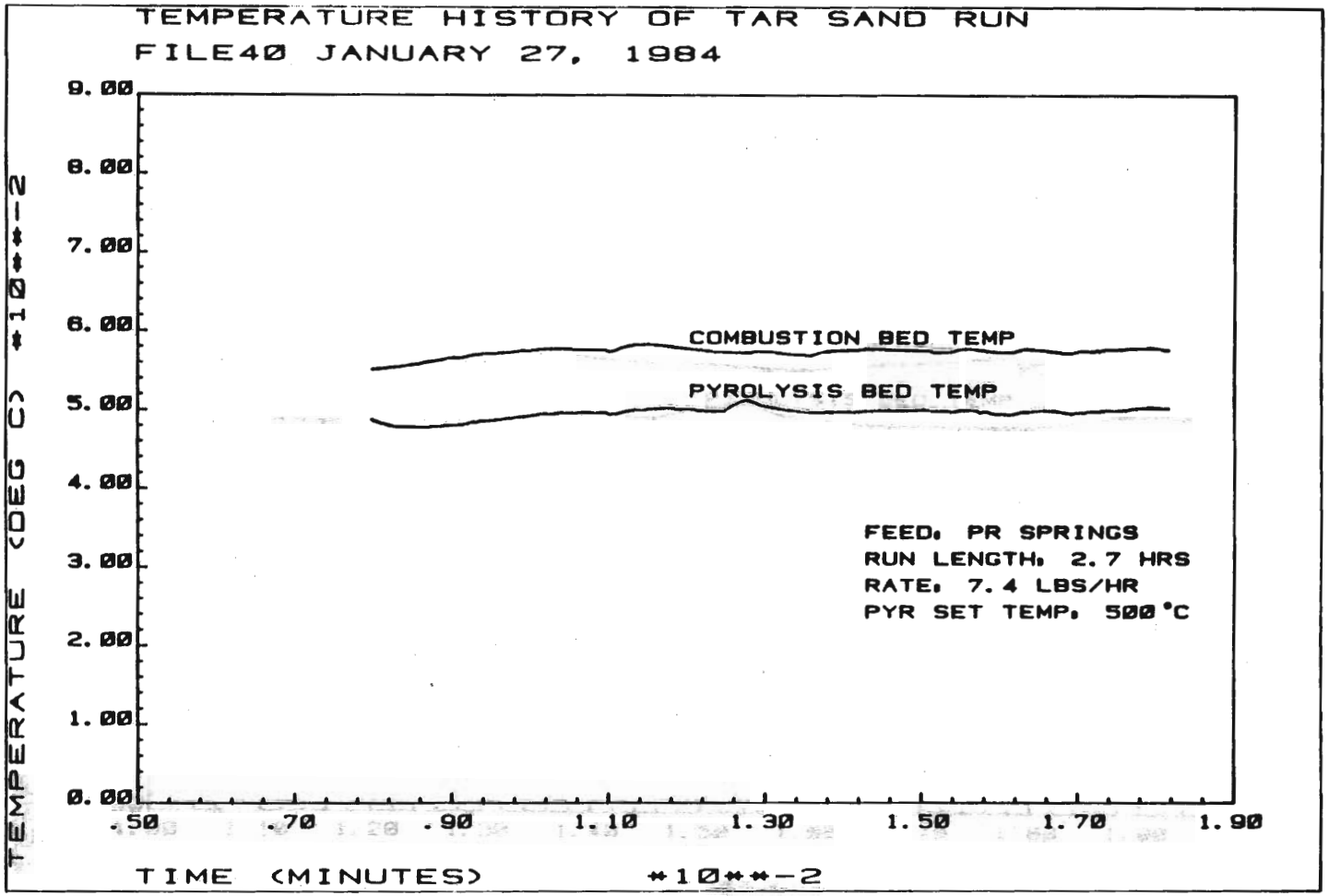

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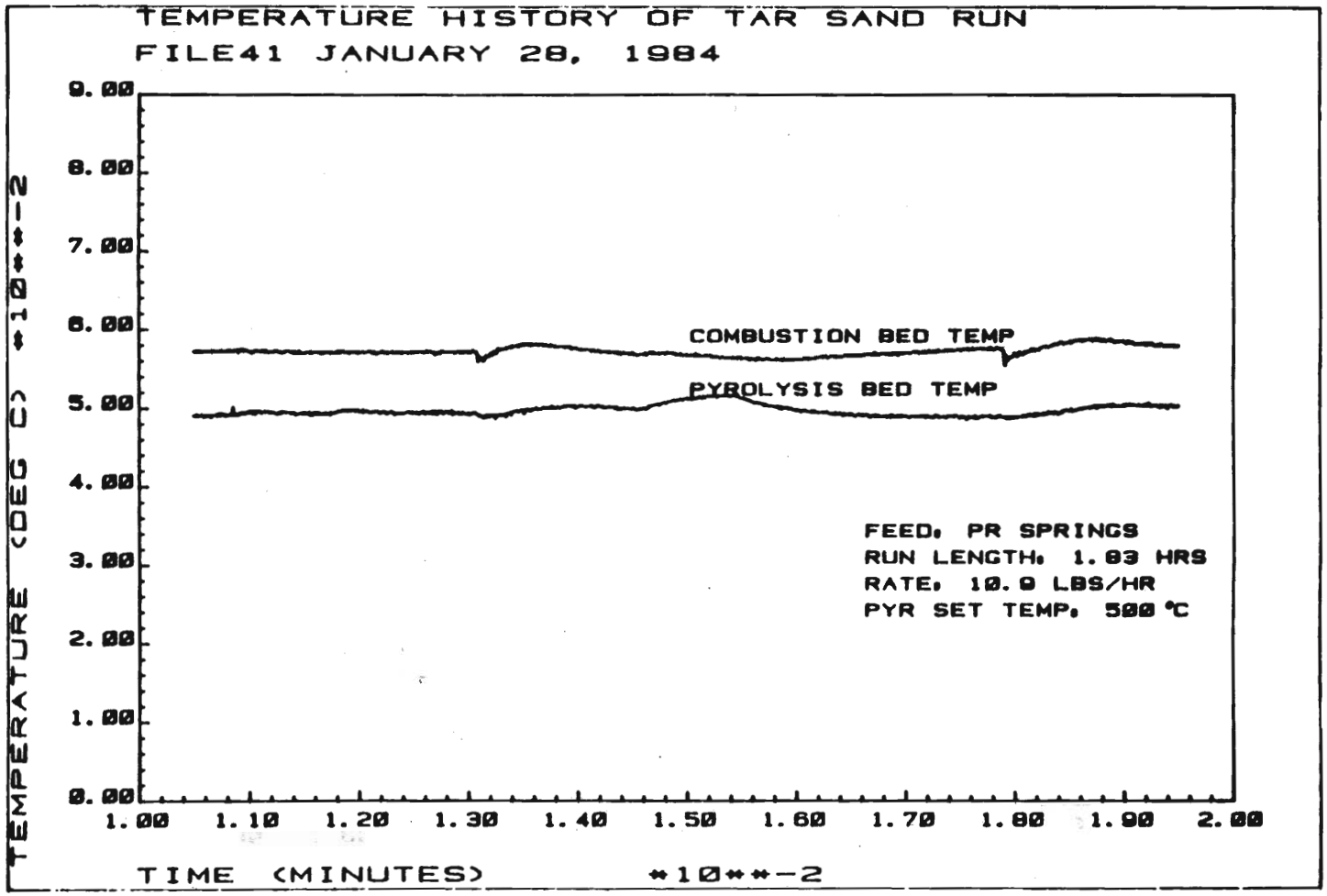

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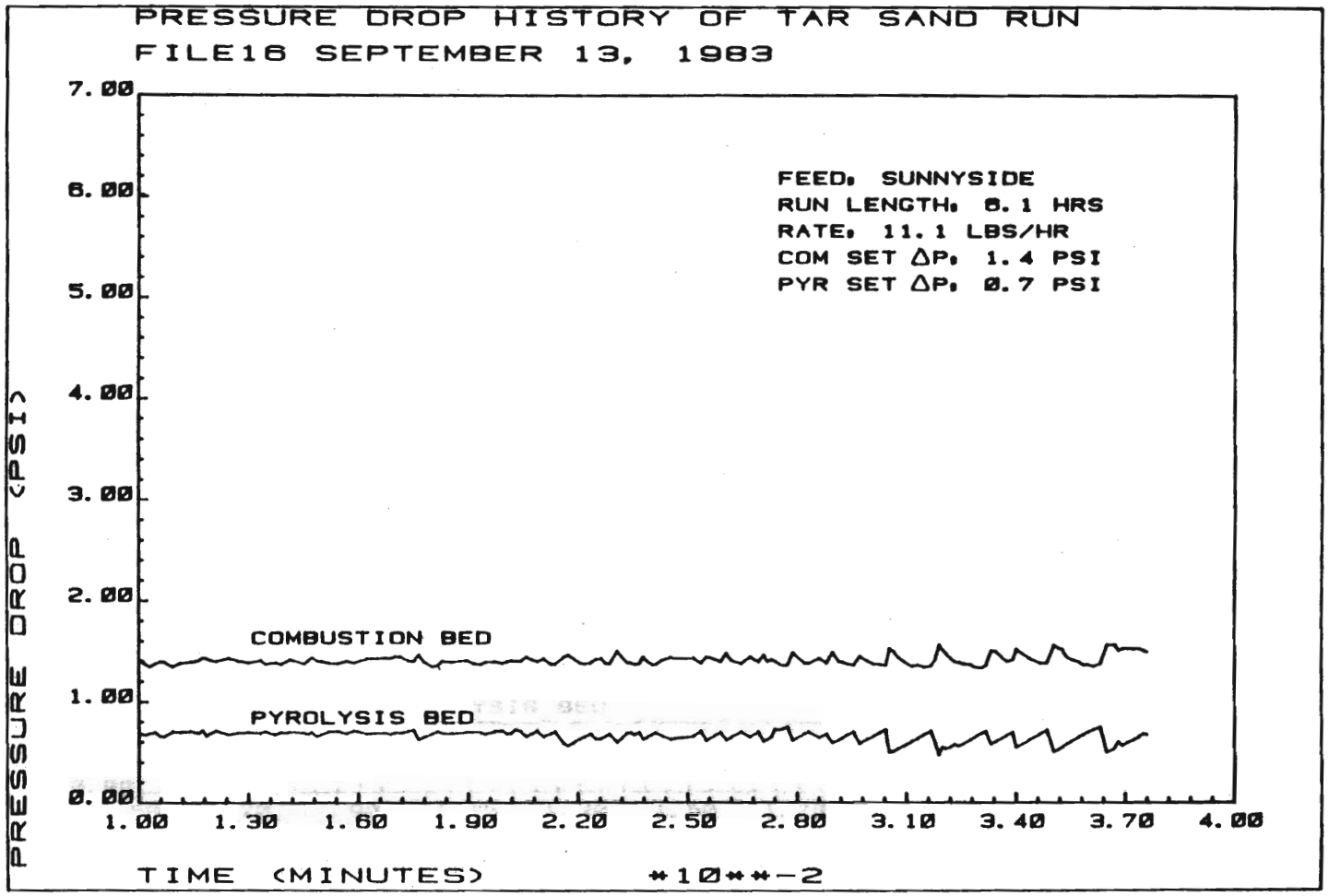
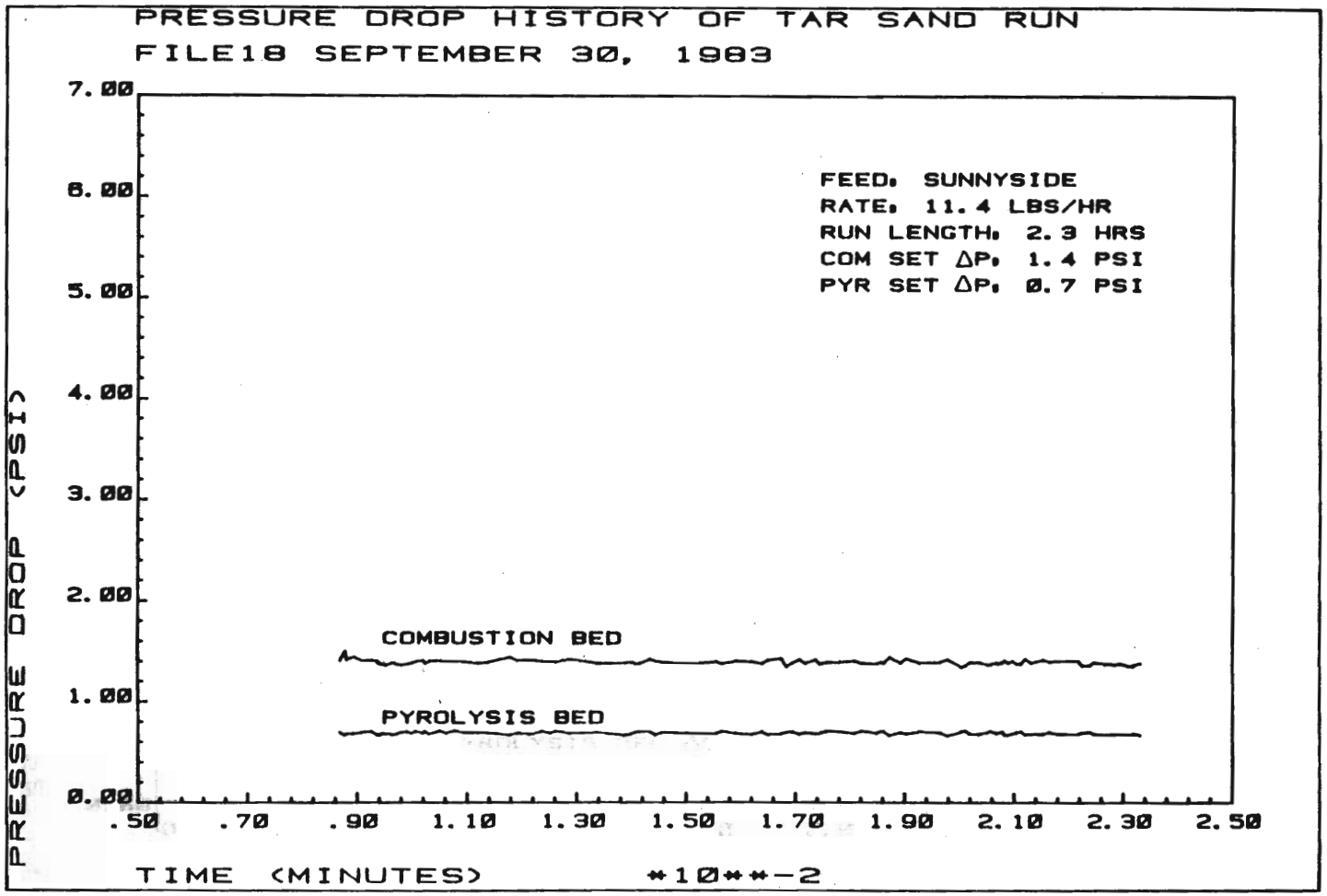


Figure 35. Pressure drop histories of tar sand run



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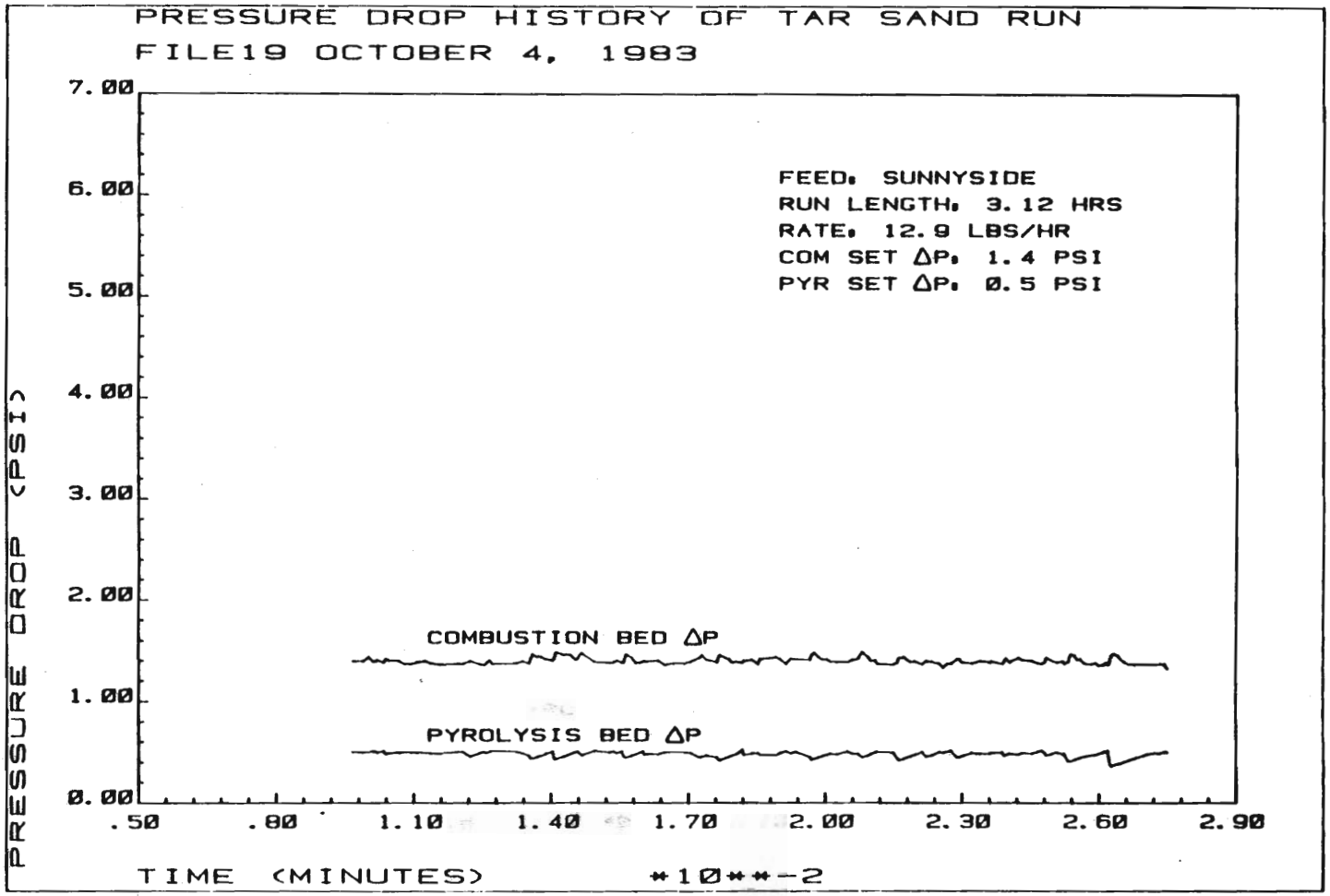

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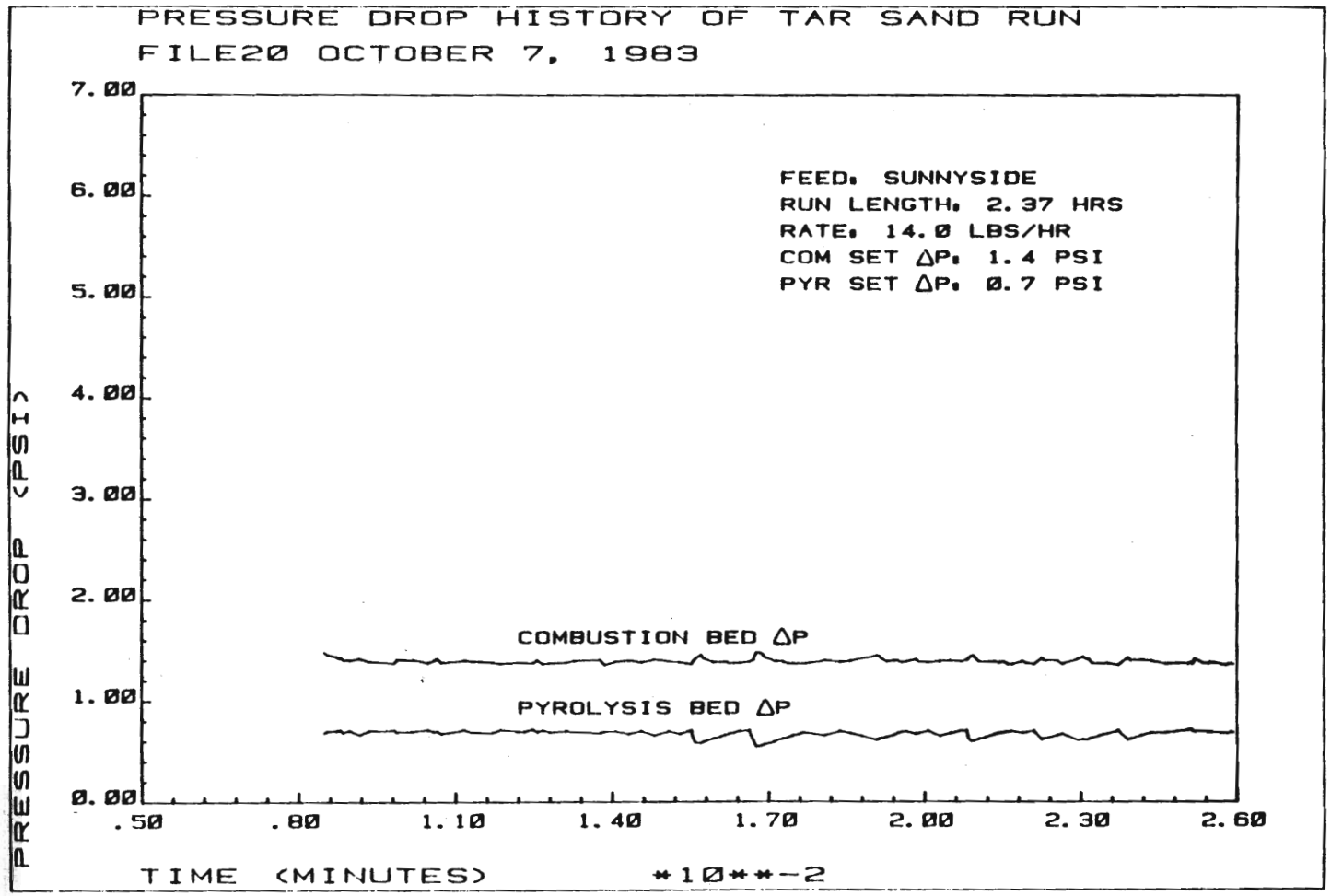

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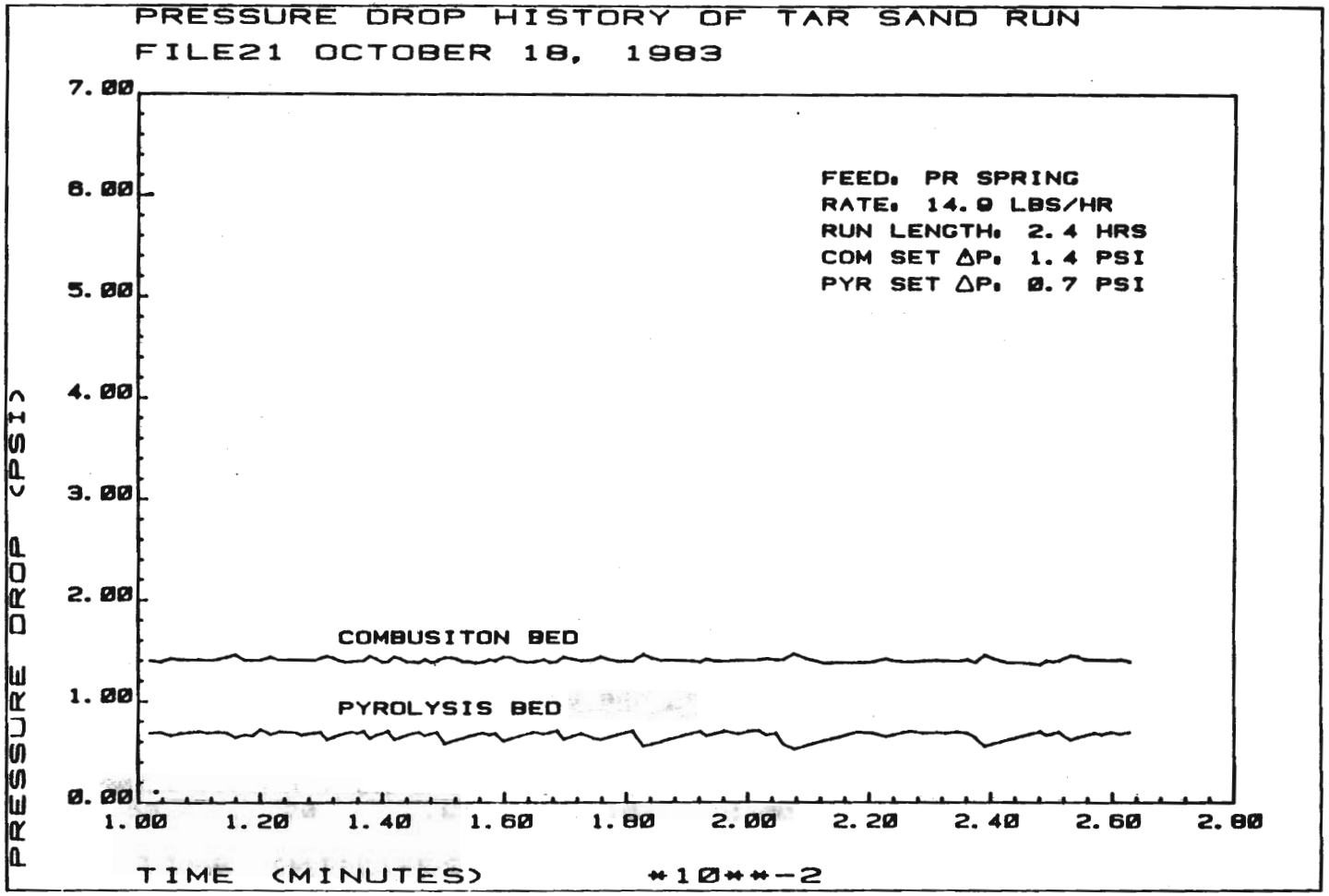

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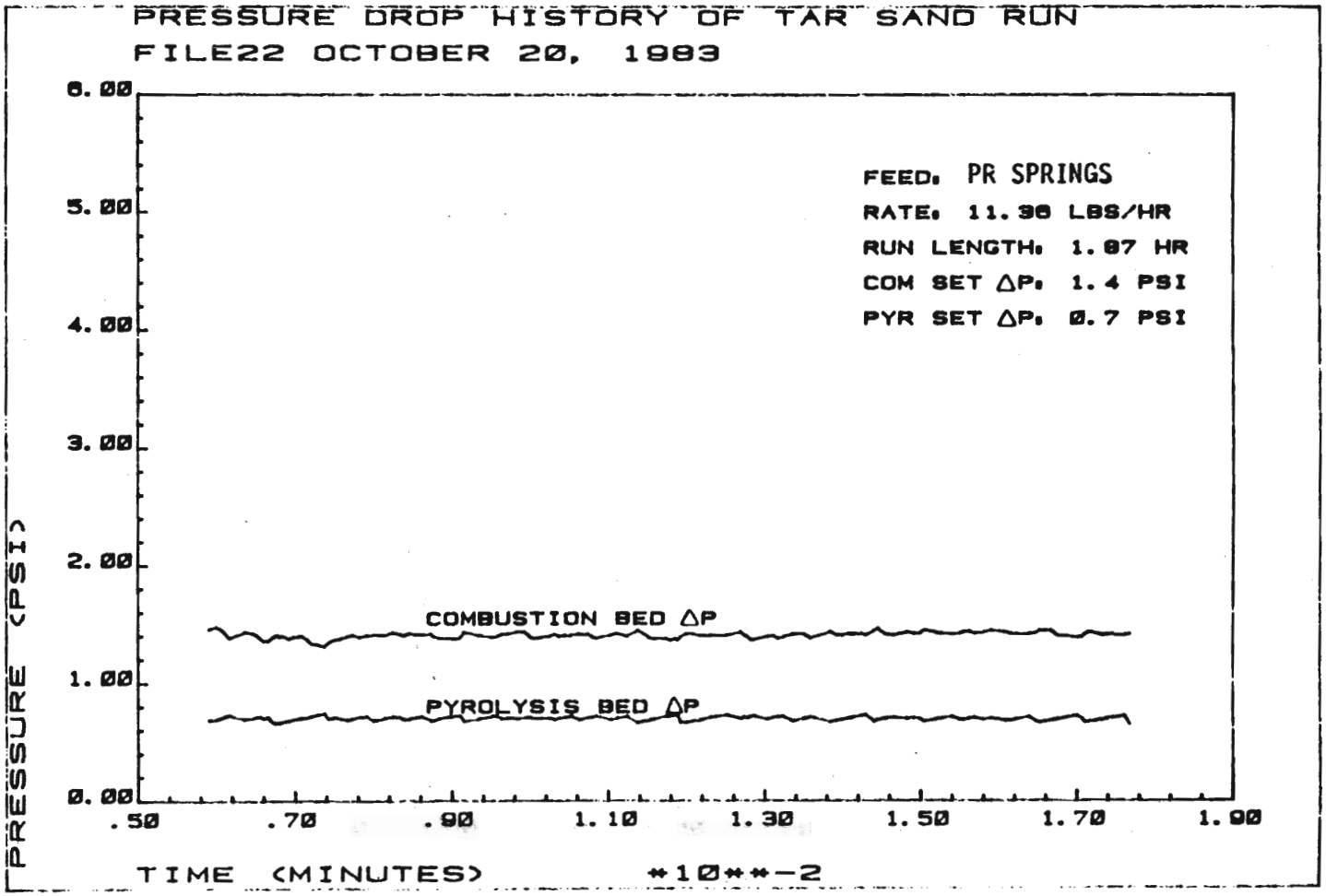

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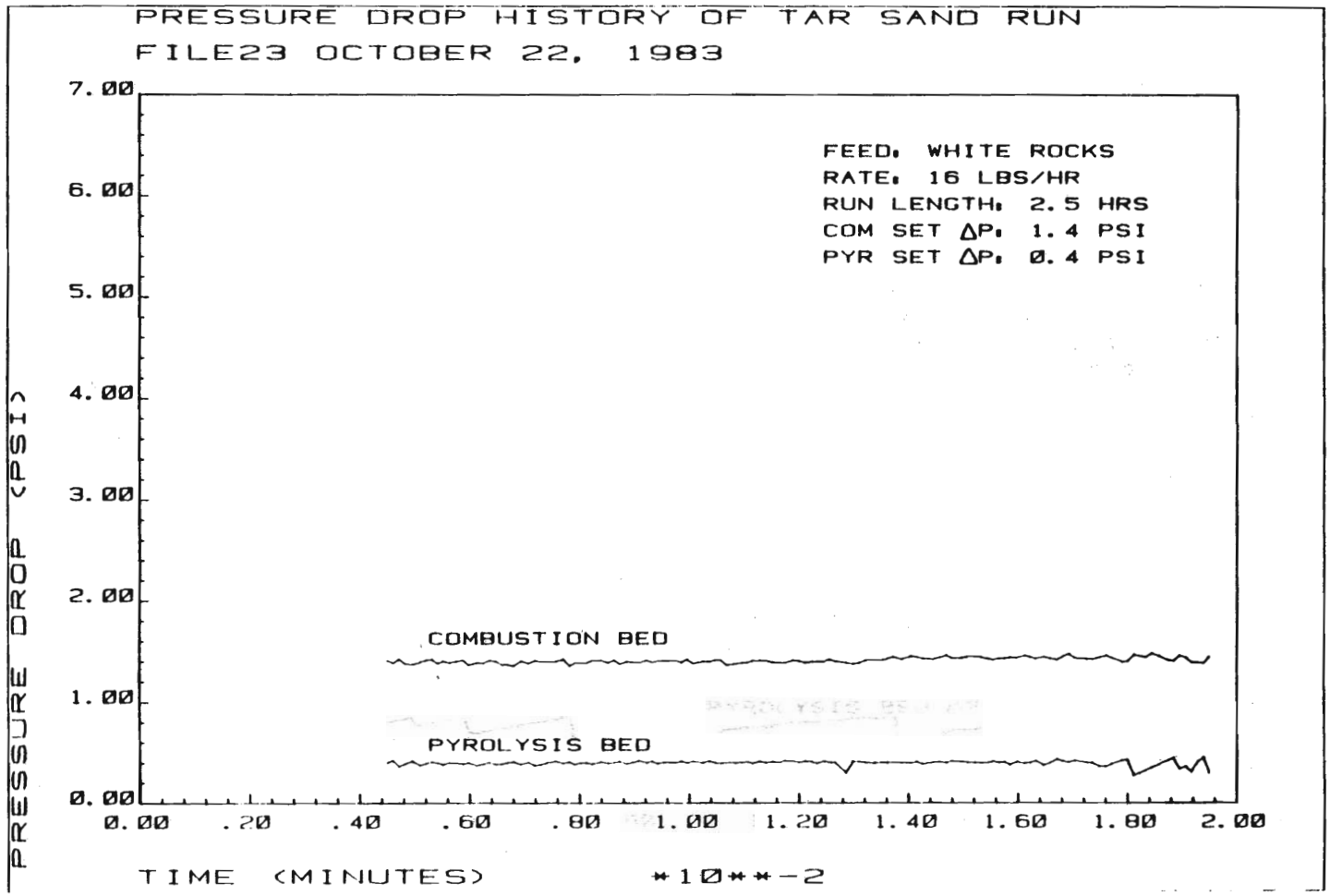

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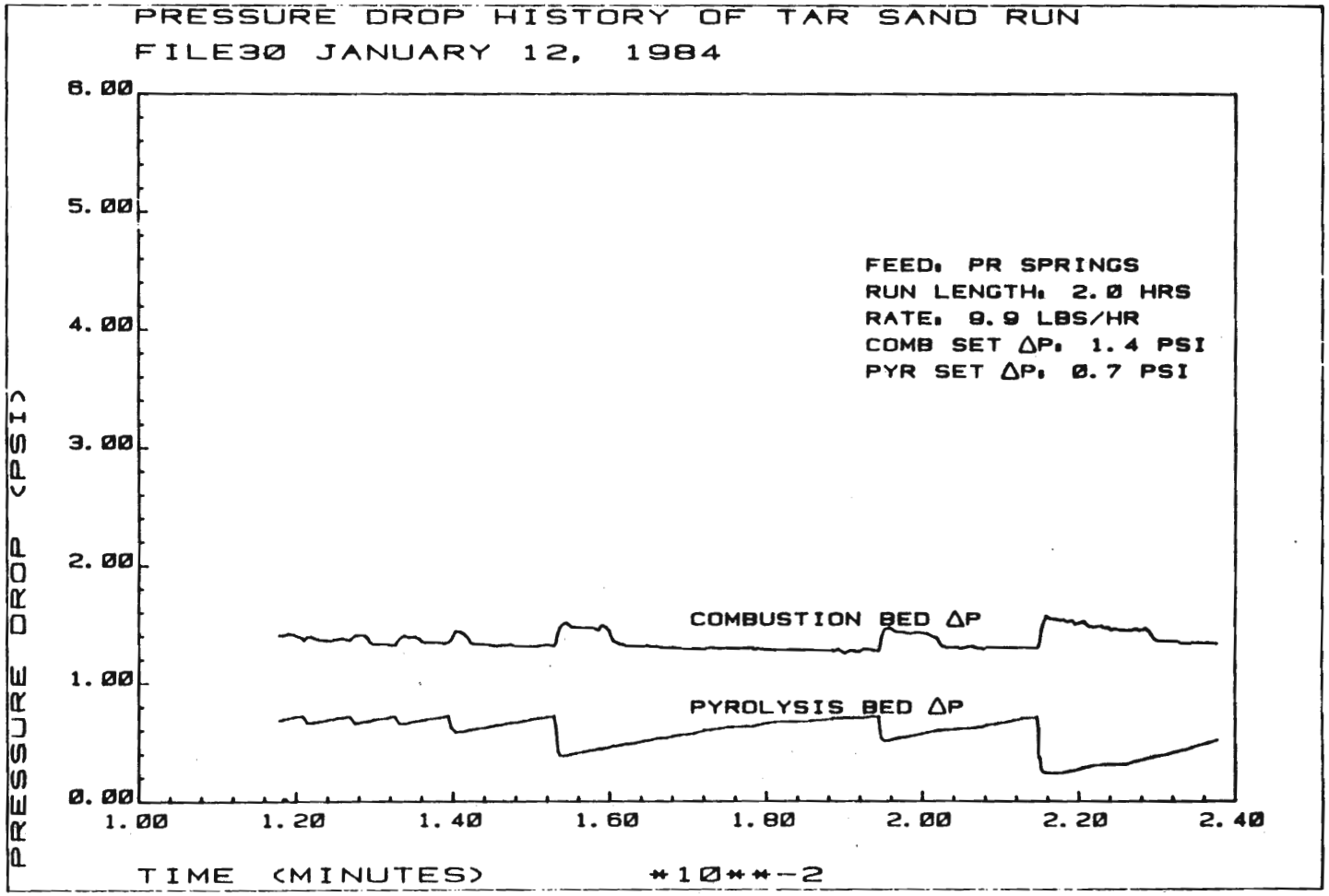

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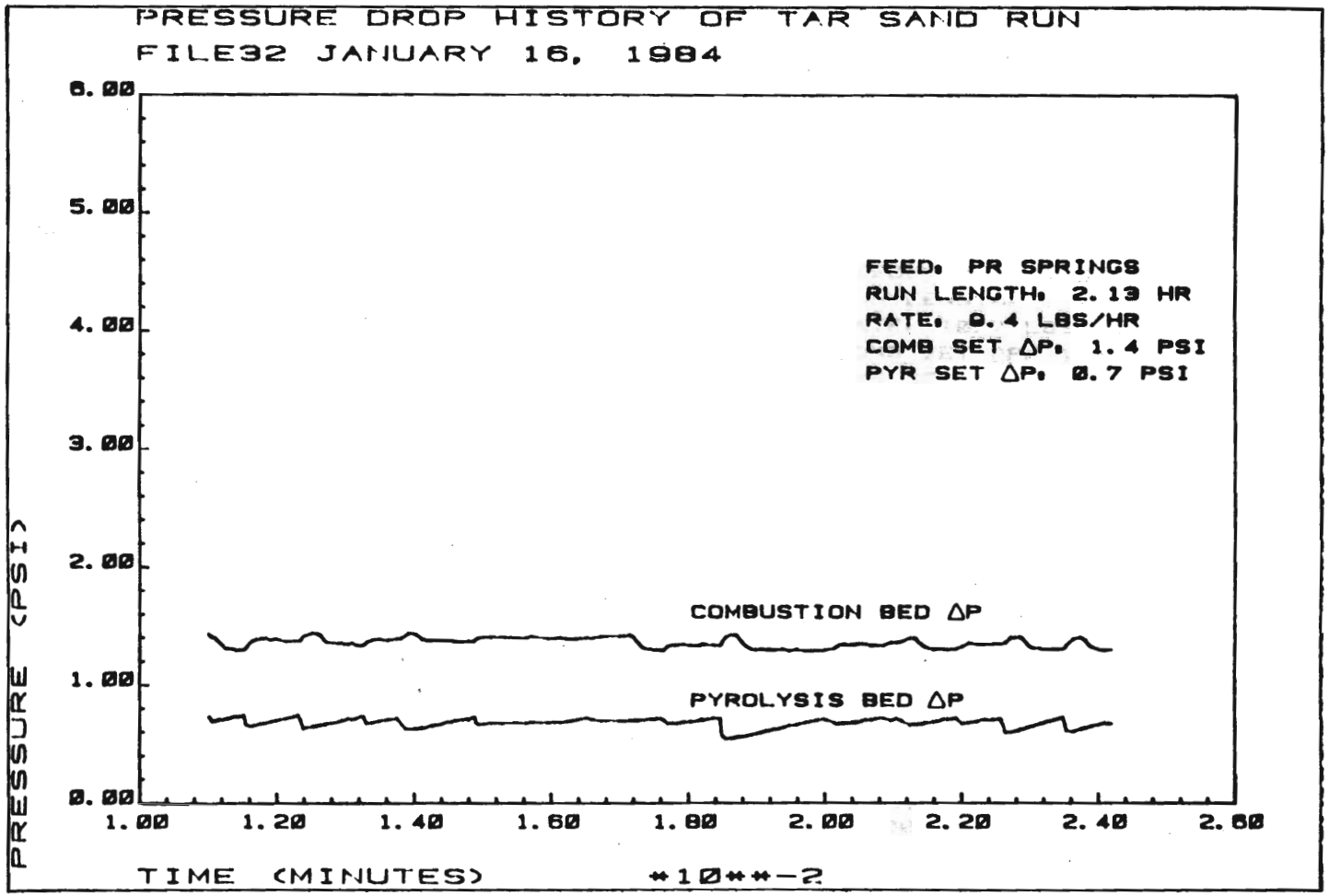

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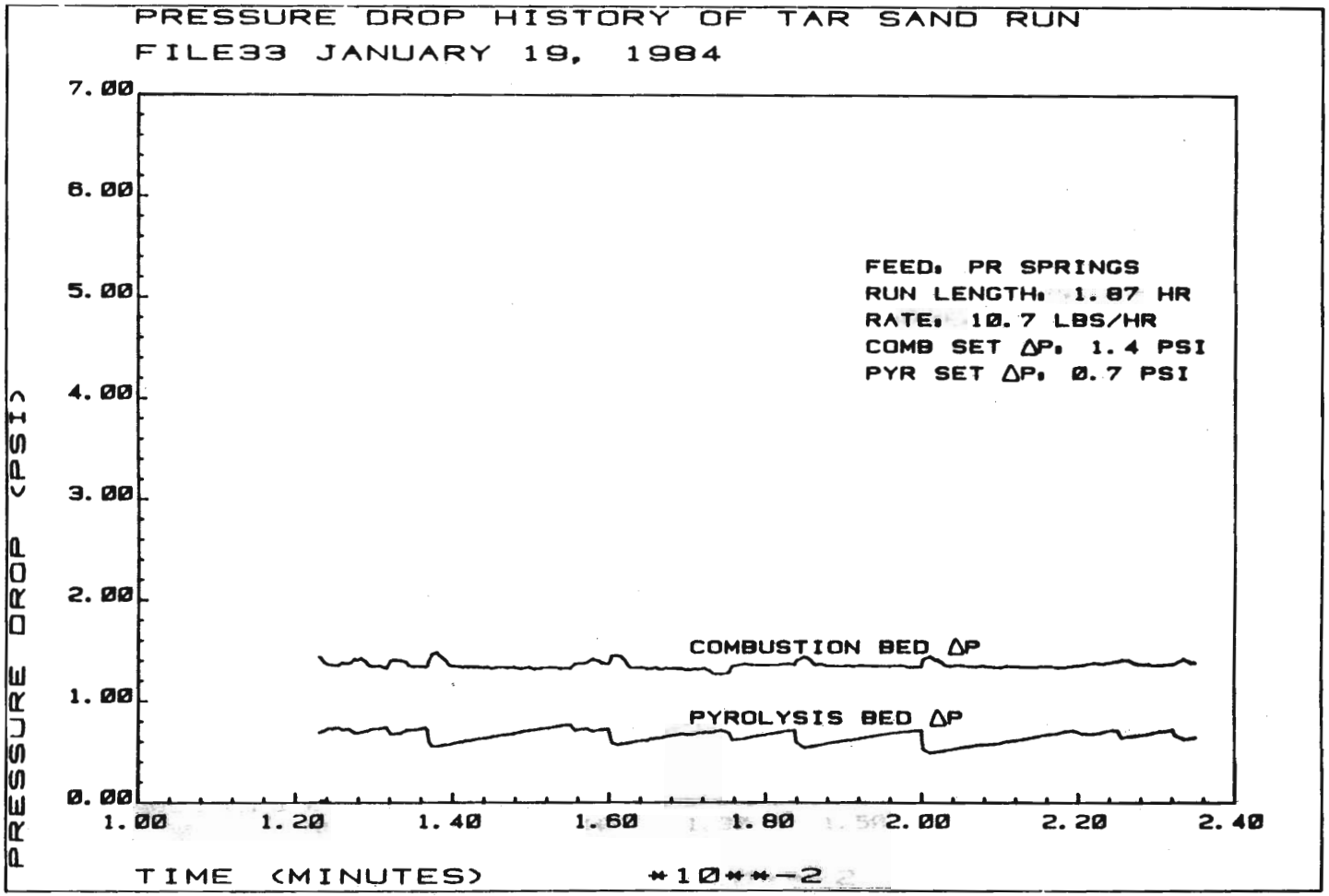

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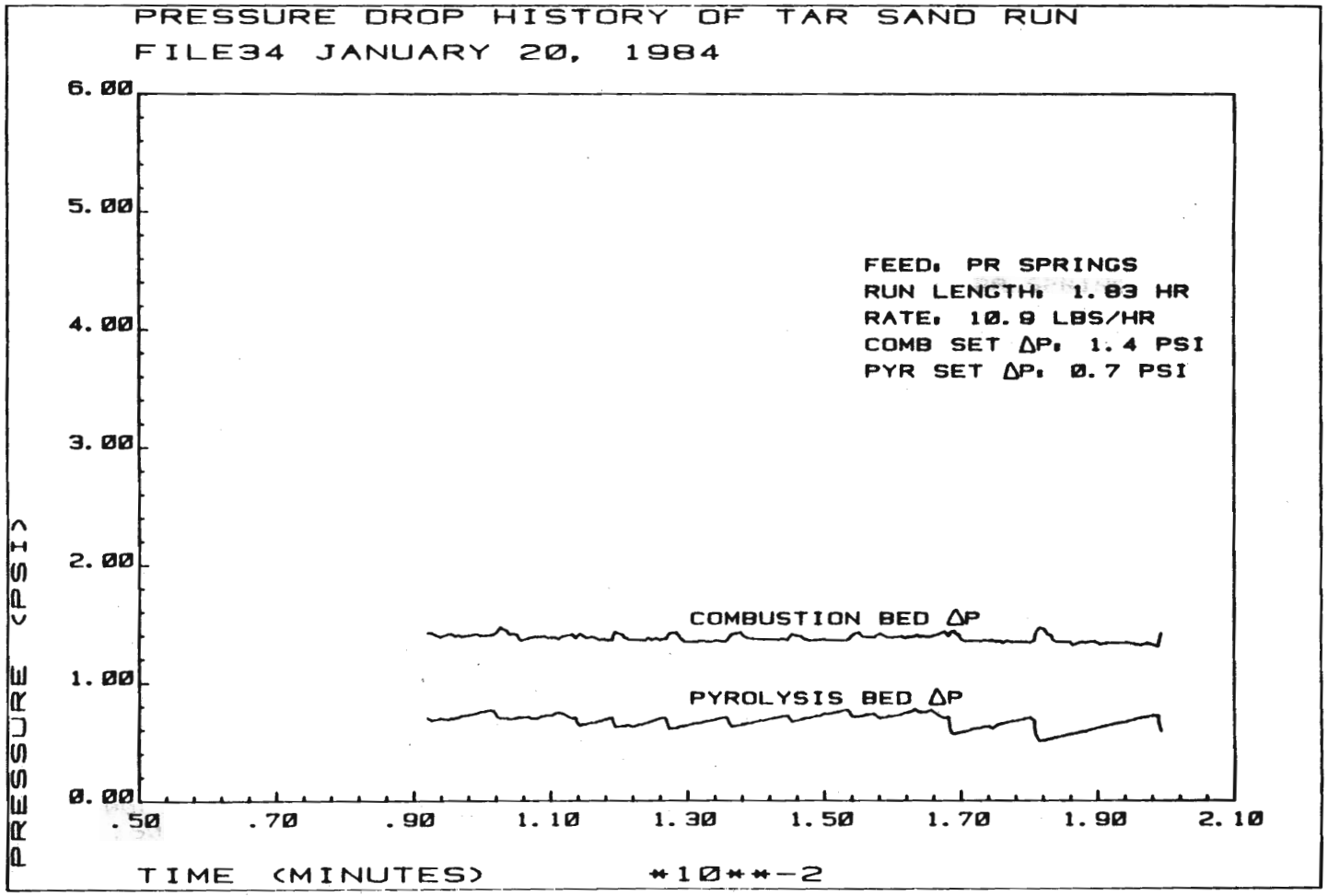
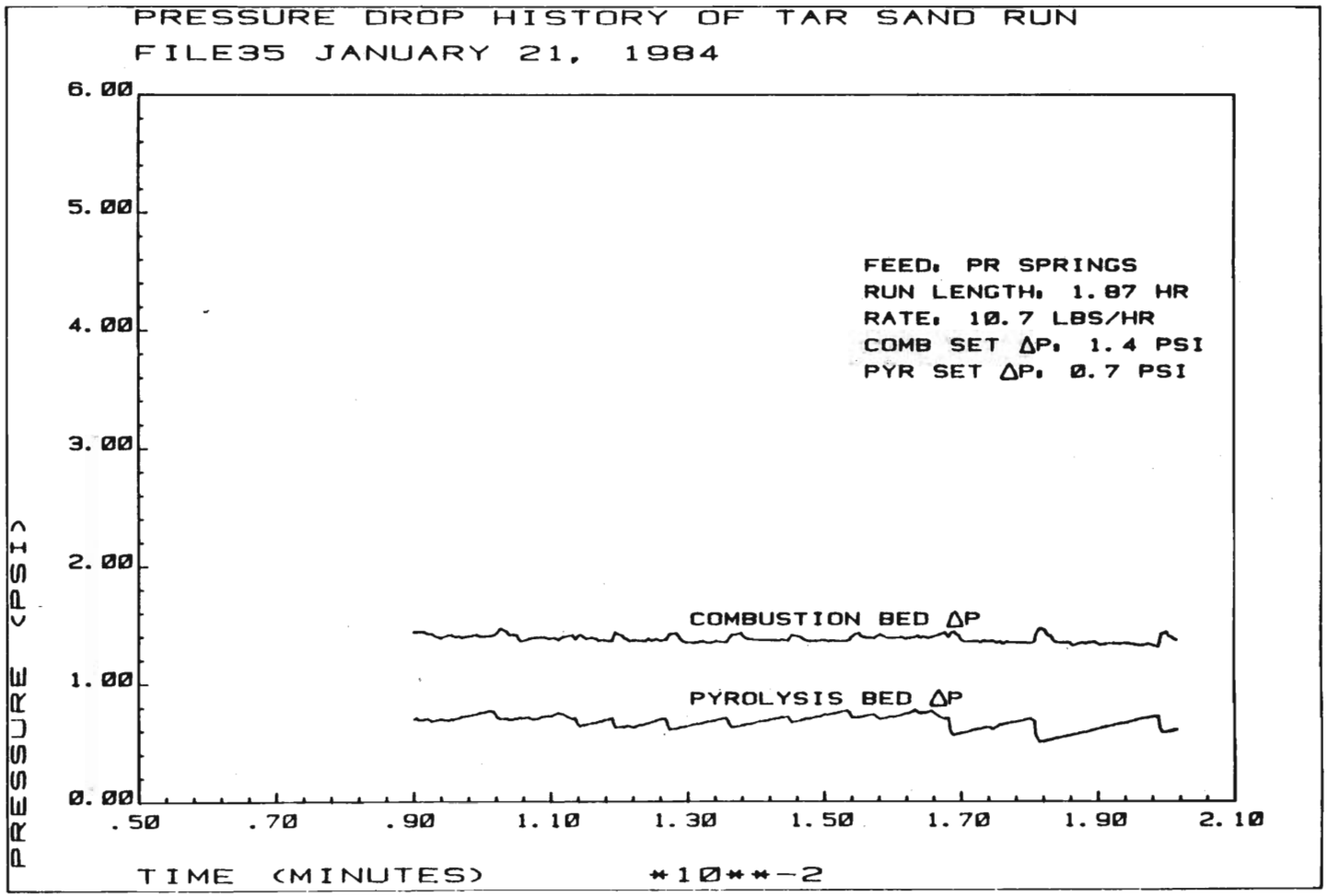


Figure 35. (continued)



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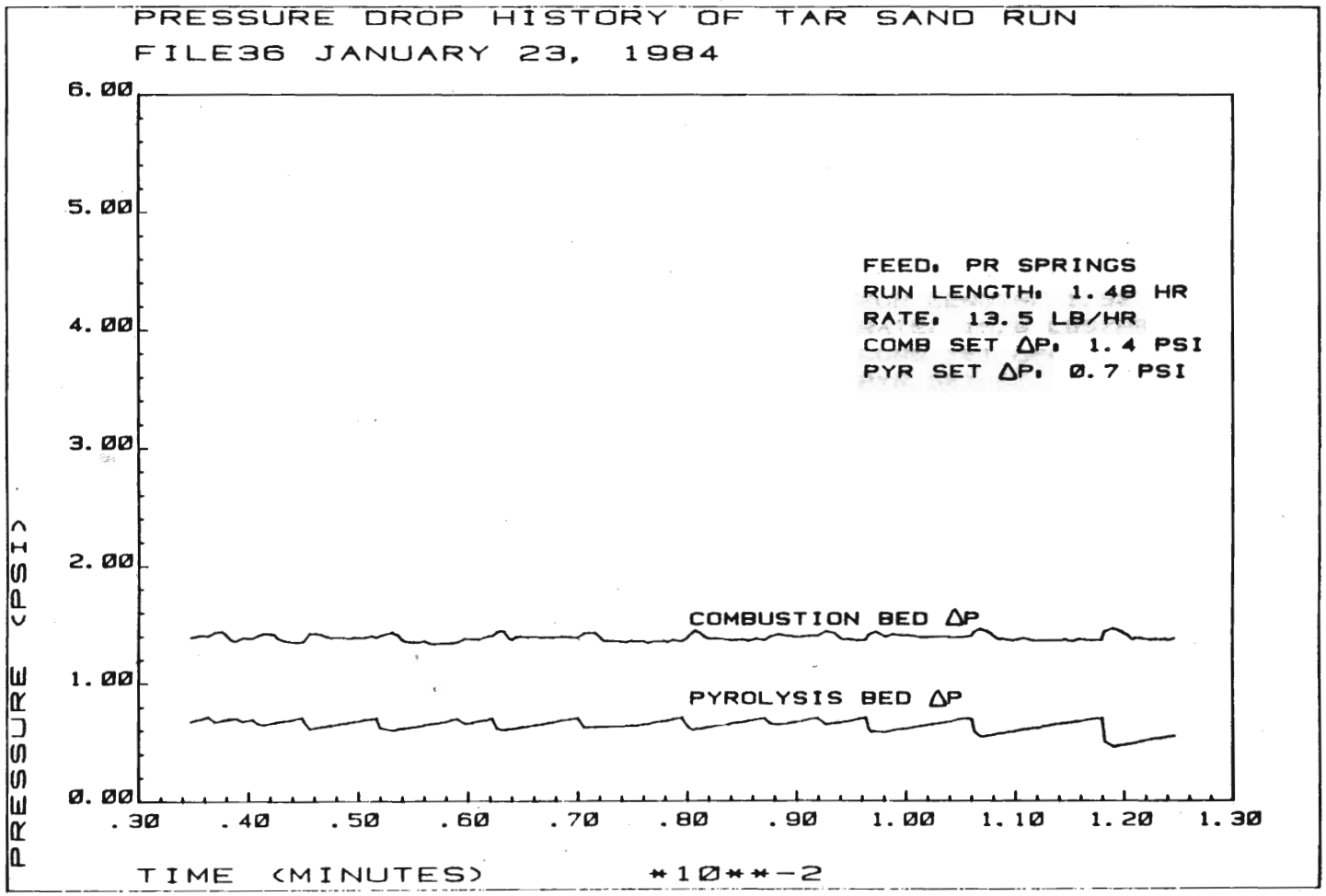

Figure 35. (continued)

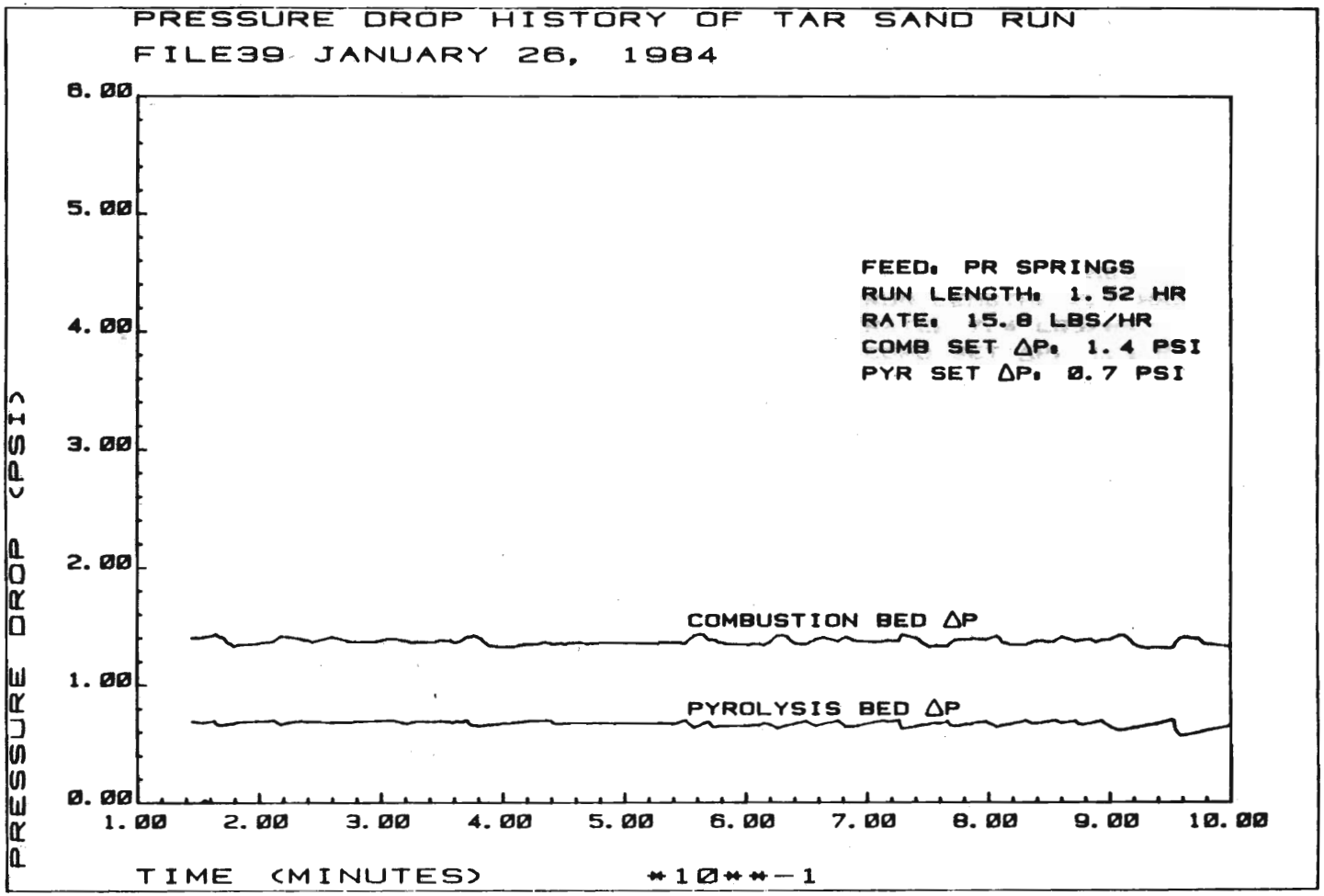

Figure 35. (continued)

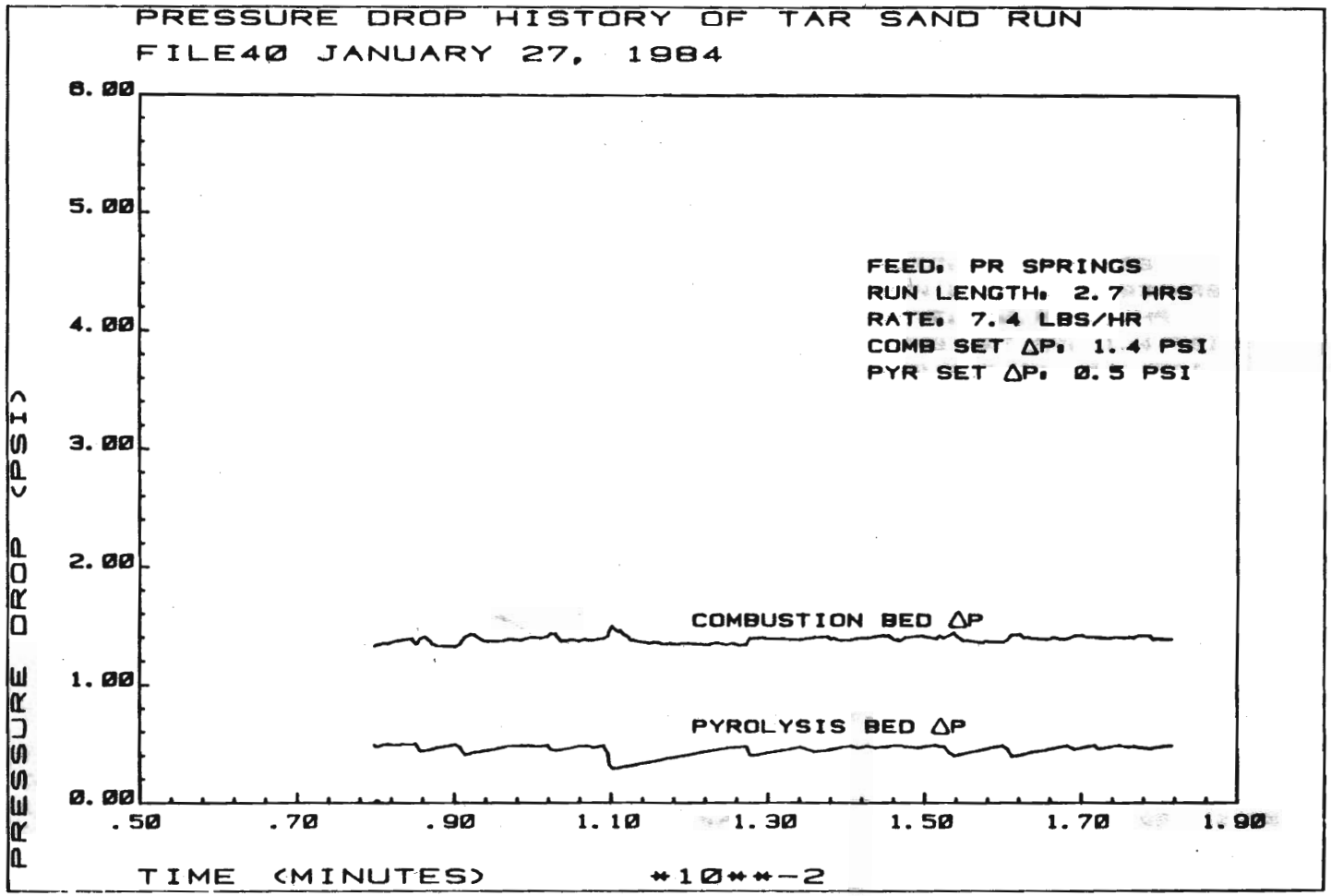

Figure 35. (continued)

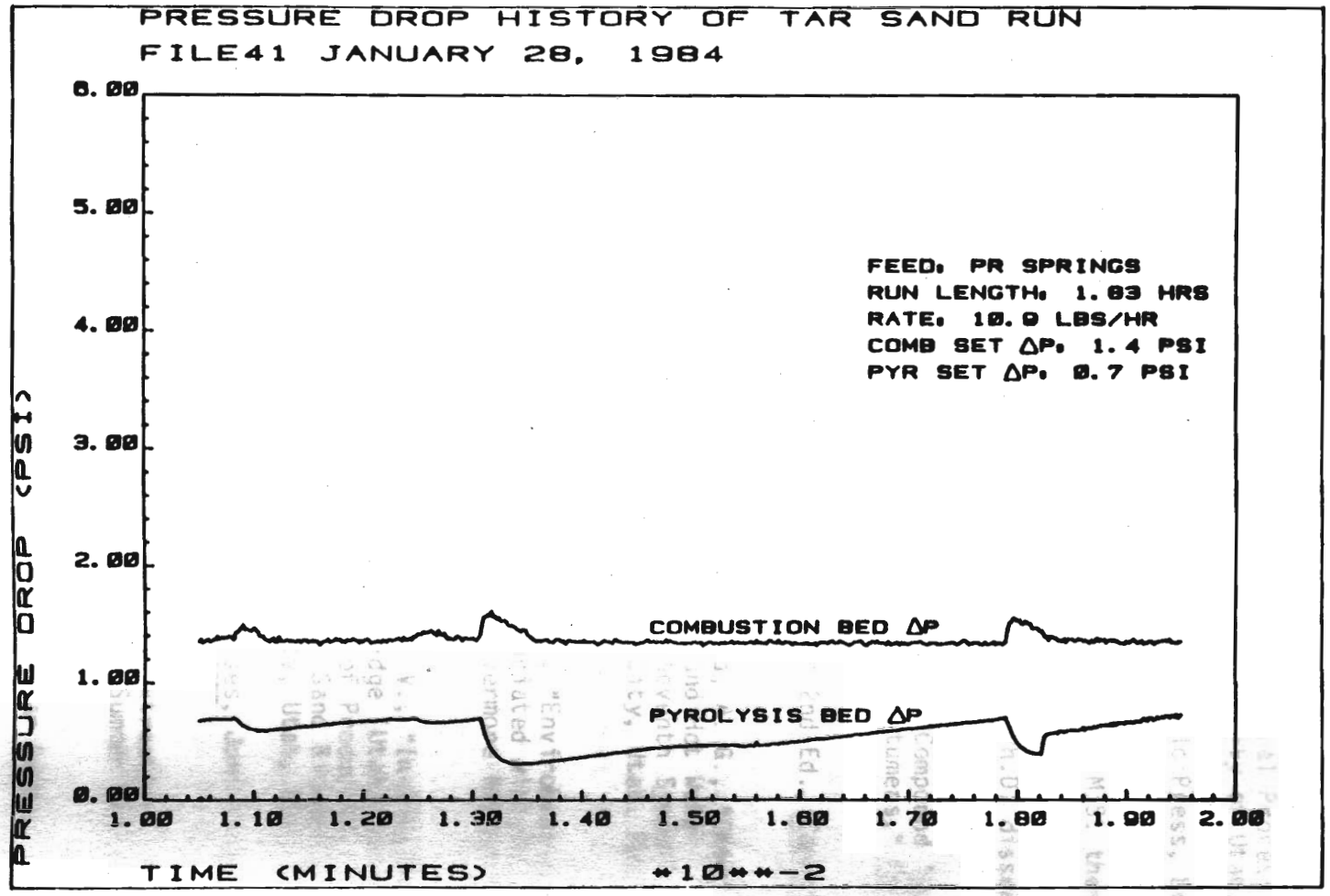

Figure 35. (continued)

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